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Air stripping of volatile organic contaminants using rotating disc contactor

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Air Stripping of Volatile Organic Contaminants Using Rotating Disc Contactor

by

Kai-Ping Hsueh

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APPROVAL OF DISSERTATION

AIR STRIPPING OF VOLATILE ORGANIC CONTAMINANTS USING ROTATING DISC CONTACTOR

BY

KAI-PING HSUEH

FOR

DEPARTMENT OF CIVIL

AND ENVIRONMENTAL ENGINEERING

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ABSTRACT

Title: Air Stripping of Volatile Organic Contaminants Using Rotating Disc Contactor

Author: Kai-Ping Hsueh, Doctor of Engineering Science, 1989

Dissertation directed by: Dr. Yeun C. Wu

Previous studies have suggested that the transfer of VOCs to the atmosphere by air stripping may be a convenient and potentially cost-effective method to remove these compounds from the water. A Rotating Disk Contactor (RDC), similarly to other aeration devices, can provide intimate contact between the VOC-bearing water and air. It is believed that under carefully selected operating conditions, the RDC system are more competitive with packed tower than other aeration methods. The objectives of this study were to evaluate the feasibility of using the RDC to strip VOCs from the contaminated water and to develop a mathematical model to predict the stripping rates. The information obtained from this study will be helpful in assessing the emission of VOCs to the atmospherefor decision makers who have to determine the appropriateness of the design and operation of a RBC system.

A laboratory-scale RDC pilot unit was used in this study to treat a synthetic wastewater containing 0.5 to 5

ppm of TCE, PCE, EB, CLF, and TCEN. Batch and continuous flow systems were applied to study the desorption of VOCs and absorption of oxygen in the RDC pilot plant. Both systems produced very similar results. The experimental data shows that the RDC pilot plant behaves like a completely mixed reactor. The mass transfer rates of the studied components were found to be a function of disk rotational speed, depth of immersion, disk radius, specific contact area of the bulk surface area and the exposed disk surface area. A semiempirical model, combining several oxygen transfer models proposed in the literature, were developed in this study. It provides a more comprehensive correlation between K_L of different compounds and the operational parameters. As indicated by the data, direct transfer from the bulk water surface dominates at rotational speeds less than 0.5 rpm. However, mass transfer through the water film on the disk surface dominates when rotational speeds are above 4 rpm.

The rate of mass transfer of VOCs is shown to be dependent on the solute's diffusivity in water to a power approximately equal to 0.5. This exponent coefficient is indepedent of the rotational speed and other operating conditions. The gas-phase resistance is negligible at high rotational speeds. The significance of the gas-phase is greatest for the less volatile components at low rotational speeds. Although this study was not designed to provide a

test of the two-resistance model, the experimental data has shown that the two-resistance approach results in a significant improvement over the simplistic assumption of liquid-phase control.

To achieve the same degree of treatment, the RDC system entails a greater volume and area than the packed tower. Since the cost per unit volume of media is lower for the RDC, preliminary economical analysis shows that the capital cost is lower for the RDC system than for the packed tower; if cheap land is available. The energy consumption is somewhat higher for the RDC system. However, the operating and maintenance labor cost is expected to be lower for the RDC system. When there is the risk of inorganic precipitation, growth of biomass, clogging by solids and particles to limit the application of packed tower, the RDC can be a good alternative for air stripping.

It has been shown that a great potential exists for the RDC system to enhance biodegrading of the VOCs with the aid of air stripping. Under carefully selected operating conditions, it is possible to decrease the VOC concentration to a level that will not violate the air pollution permit and at the same time, allow biodegradation to occur. The combination of the air stripping process with

the biological process, in a RDC system to remove VOCs deserves further study and evaluation for the control of VOCs.

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NOTATION

At	Surface area of the bulk water, m^2
Aa	Exposed disk area, m^2
Ap	Projected media area on the water surface, m^2
a	Specific contact area, 1/m
a_l	Specific contact area of water surface, 1/m
a_d	Specific contact area of disk surface, 1/m
at	Empirical constant
a'	Empirical constant
b'	Empirical constant
C	Concentration of solute, mg/l, or the empirical constant in Eq. 5.1
$C_{a,t}$	Concentration in the water film at the point of immersion at time t, mg/l
Ca^*	Effective average solute concentration, mg/l
Ci	Influent concentration (in continuous flow system), or initial concentration (in batch system), mg/l
$C_{i,t}$	Concentration in the water film at the point of emergence at time t, mg/l
Ce	Effluent concentration, mg/l
Cs	Equilibrium concentration, mg/l
Ct	Concentration at time t, mg/l
D	Diffusivity of the solute in water, m^2/sec
Di	Diffusivity of a specific VOC
Do	Diffusivity of oxygen
f	Overall interchange ratio coefficient

f1	Interchange ratio coefficient
f2	Interchange ratio coefficient
fv	Void volume, m^3
G	Volumetric gas flow rate, m^3/sec
g	Acceleration of gravity, m/sec
H	Depth of immersion, m
Hc	Henry's constant (dimensionless)
K	Proportionality constant for power dependence
K'	Empirical constant
K1	Empirical constant, 5.7×10^{-8}
K2	Conversion factor, 35.32 mg ft^3/g l
K_L	Overall mass transfer coefficient, m/sec
K_{Ll}	Overall mass transfer coefficient for water surface, m/sec
K_{Ld}	Overall mass transfer coefficient for disk surface, m/sec
K_La	Overall mass transfer rate constant, m/sec
K_La_d	Overall mass transfer rate constant for water film transfer, m/sec
K_La_l	Overall mass transfer rate constant for water surface transfer, m/sec
K_La_i	Overall mass transfer rate constant for VOC
K_La_i	Overall mass transfer rate constant for oxygen
k_L	Liquid phase mass transfer coefficient, m/sec
$\bar{k}_L$	Mean liquid phase mass transfer coefficient, m/sec
k_G	Gas phase mass transfer coefficient, m/sec
L	Volumetric water flow rate, m^3/sec
m	Area of the disk/ area of circle with the same dia.
Nt	Mass flux at time t, kg/m^2 sec

$\bar{N}t$	Average mass flux, kg/m^2 sec
nt	Mass flux at time t for elememt dA, kg/m^2 sec
Nv	Renewal number
n'	Number of disks
P	Power input, KW
p	Wetted perimeter, m
Q	Water flow rate, m^3/sec
q	Number of disk per length
R	Radius of the disks, m
Re	Reynold numbr
Ri	Distance between water level and shaft, m
R1	Stripping factor
Rh	Dimensionless immersion depth = H/R
R_G	Gas-phase resistance, sec/m
R_L	Liquid-phase resistance, sec/m
r	Distance on the disks to the central shaft,m
S	Half-space between disks, cm
T	Time for each rotation, sec or min
t	Time, sec
ta	Exposure time (or contact time), sec or min
$\bar{t}a$	Mean exposure time (mean contact time), sec or min
tw	Submerged time, sec or min
U	Superficial velocity, m/sec
V	Volume of the reactor, m^3
Vi	Volume of a single stage, m^3
v	Withdraw velocity of the disk, m/sec

Vc	the vertical component of the disks' rotational speed, m/sec
w	Rotational Speed (angular velocity), rad/sec or rpm
Xi	Influent mole fraction
Xe	Effluent mole fraction
x	Distance in film layer, m
Z	Packing height (of packed tower), m
α	Empirical constant, m
α'	Empirical constant
ϕ	arccos (Ri/R)
φ	arccos (Ri/r)
δ'	Thickness of the water film
δ	Thickness of the disk surface boundary film
ν	Kinematic viscosity of water, m^2/sec
μ	Dynamic viscosity of water, N sec/m^2
ζ	Dimensionless radius = r/R

CHAPTER 1 INTRODUCTION

Recent sampling and analysis of surface and ground waters throughout the country have indicated that many have been contaminated by various organic chemicals. Of all of the compounds detected, volatile synthetic organic chemicals (VOCs) have been some of the compounds observed most frequently (Council on Environment Quality, 1981;). They are named VOCs because of their distinctive common property of high volatility relative to other organic substances such as phenols, pesticides, or PCBs. Although many compounds are volatile, the VOCs mentioned in this study will be the 31 purgeable compounds on the EPA's list of 129 priority pollutants. Most of them are widely employed in various industrial, commercial, agricultural, and household activities, and considered harmful or causing potential health risk to continually exposed human beings. In a 1977 report to the Congress on Disposal Practices and Their Effects on Ground Water EPA identified the disposal of industrial wastes at industrial impoundments and solid disposal sites as the major source of groundwater contamination. The secondary sources are from land disposal of municipal wastes, discharges of treated and untreated wastewater, and activities related to mining, agriculture and petroleum exploration and operation (EPA, 1977).

Table 1.1 summarizes the names of the 31 volatile organic chemicals and the relative frequency with which they are being found in industrial wastewaters. This data was obtained from the results of EPA's investigation on 31 August 1978. A total of 32 industrial categories and subcategories were analyzed for organics (Keith, 1979). The percent of samples in this table represents the number of samples in which the compound was found, divided by the total number of samples. According to the relative frequencies and magnitudes of occurrence of the volatile organic priority pollutants in surface water, groundwater, and wastewaters and their effects on human health and aquatic life, five of the compounds listed in Table 1.1 were chosen as target compounds in this study. They were trichloroethylene (TCE), tetrachloroethylene (PCE), 1,1,1-trichloroethane (TCEN), chloroform (CLF), and ethylbenzene (EB). These five compounds are frequently observed in surface water, groundwater and in wastewaters. Four of them, i.e. TCE, PCE, CLF, and TCEN, are confirmed human or animal carcinogens (Petura, 1981). In addition to their carcinogenicities, all the five VOCs cause acute and chronic effects on the central nervous system, the lungs, the liver, and the kidneys with common symptoms as headaches, dizziness, vertigo, tremors, nausea and vomiting, irregular heart beat, sloppiness, fatigue, blurred vision, coma, and intoxication similar to that from alcohol (NIOSH, 1985).

Most of the VOCs are used by industries as solvents. For example, trichloroethylene is mainly used as a degreasing solvent in metal industries, and tetrachloroethylene as a textile scouring solvent and metal degreaser. The occurrence and concentrations of the five VOCs in different industrial wastewaters are listed in Table 1.2 to 1.6. They are found most often in metal finishing, pulp and paperboard mills, and textile mills. Their concentrations vary largely with different industries. The maximum concentration may reach 130 ppm for TCE, 110 ppm for PCE, 280 ppm for chloroform, 1300 ppm for 1,1,1-Trichloroethane, and 42 ppm for ethylbenzene (EPA, 1983). The occurrence of these volatile priority pollutants in municipal wastewater were investigated by EPA in Publicly Owned Treatment Works (POTWs). The results showed that the contribution of VOCs to a POTW's influent was associated largely with specific industrial categories, and that only a very small part of VOCs came from commercial and household activities like rug cleaning, drain cleaning, etc. The detected frequency and concentration range of the 5 selected VOCs in POTW influents and effluents are summarized in Table 1.7 (EPA, 1982).

The presence of VOCs in drinking water supplies is generally believed to be caused by improper waste disposal practices. A considerable data base has now been establis-

hed for VOCs in selected surface water supplies. It shows that most supplies contained relatively low concentrations of the VOCs. Table 1.8 adopted from "The Occurrence of Volatile Synthetic Organic Chemicals in Drinking Water" (EPA, 1981) shows that most of these compounds were found with concentrations less than 10 ppb. Mean VOC concentrations ranged from 0.5 ppb to 3.5 ppb while maximum values laid in the range of 10 times the average values. This data provides an overview of the national situation but do not address the infrequent, short-term occurrences of high concentrations resulting from spills and deliberate discharges. Another program managed by the Ohio River Sanitary Commission (ORSANCO, 1981) found that chloroform is prevalent throughout the river basin at very low concentrations. The next most commonly found compounds were 1,1,1-trichloroethane and tetrachloroethylene. Similar to the data of EPA's survey, the VOCs usually occur in very low concentrations, typically less than 1.0 ppb.

Generally, contaminated surface waters have been found to contain a broad spectrum of organic compounds at very low concentrations. Contaminated groundwaters, however, usually contain a lesser number of compounds but they occur at much higher concentrations. The property of volatility and larger Henry's constants are potential reasons for the low VOCs' concentrations in surface water. However, for

groundwater flow, the limited contact with the atmosphere and limited microbial activity prevents the quick depletion of VOCs. Once VOCs are introduced into a groundwater system, the aquifer tends to preserve more than dissipate them (Miller, 1985).

Accordingly, several groundwater surveys by EPA, like the National Organic Reconnaissance Survey (NORS), National Organic Monitoring Survey (NOMS), Community WAter Supply Survey (CWSS), and the Ground Water Supply Survey (GWSS), found that volatile chlorinated hydrocarbons were frequently present, and sometimes in very high concentrations (EPA, 1981). Table 1.9, 1.10 show the results of these surveys. The maximum concentration of 1,1,1 trichloroethane can be as high as 400 ppm, and trichloroethylene may reach 35 ppm. The concentrations of the selected VOCs, according to the Council on Environmental Quality's (CEQ) survey in 34 states, are summarized in Table 1.11.

The widespread contamination of waters by VOCs has resulted in an awareness of the need for techniques of removal of such pollutants. Various methods of eliminating VOCs from the source of public drinking water have been investigated. (Guswa, et. al., 1984; Wilson, 1983; Singhal 1981). They basically fall into two categories: management techniques (i.e., find new source of water supply or eliminate the source of contamination) and treatment

techniques. Previous studies on treatment techniques indicated that adsorption processes using activated carbon or adsorbent resins, chemical oxidation with ozone, or chlorine dioxide, and reverse osmosis are capable of removing VOCs from the contaminated source of potable water supply but these treatment alternatives are very expensive and operationally complex.

A number of VOCs, such as benzene, ethylbenzene, methylene chloride, and toluene may be removed through biodegradation (Namkung, 1987). However, biological treatment of low molecular weight VOCs, such as trichloroethylene, tetrachloroethylene, trichloroethane, and chloroform etc., were found to be difficult to treat in either aerobic or anaerobic environments and, therefore, the treatment efficiency was low (National Research Council, 1984; Roberts, et al. 1980; Nelson, 1986; Wilson, 1985). The low concentrations of VOCs usually found in wastewater may retard an induced enzyme formation by the microoganism and the formation of an acclimated microbial population to degrade the contaminants. The high volatilization losses of VOCs in biological processes may further decrease the concentration thereby, hindering the biodegradation. With this point of view, a biological process should minimize the volatilization loss as much as possible.

Previous studies have suggested that the transfer of VOCs to the atmosphere by air stripping may be a convenient and potentially cost-effective method to remove these compounds from the low contaminated water, e.g. drinking water, surface water, ground water (Smith, 1980; Symons, 1981, Singley, 1983, Roberts, 1983). Several aeration methods have been investigated or applied for this purpose: mechanical surface aeration, diffused aeration, spray or tray towers, spray fountains, and countercurrent packed towers. Among them, the countercurrent packed tower has recently received the most considerable attention. Although air stripping is difficult to optimize for waters containing a spectrum of volatile and nonvolatile components, the process is frequently selected for the removal of VOCs from contaminated ground water or for the control of THM in drinking water. In addition, it does have appeal as pretreatment prior to another process, such as adsorption, in order to extend the life of the sorbent by removing sorbable VOCs. At present, the application of packed towers is still limited in treating low particle-bearing waters, such as ground water or drinking water. Most packed towers experienced operational problems associated with precipitation of inorganics or growth of biological slime. Due to the high particles and organic content in industrial wastewater, these plugging problems are expected to be more severe.

Low-cost Rotating Disk Contactor (RDC) biotechnology, better known as rotating biodisk contactor (RBC) has been widely used for treatment of municipal and industrial wastewaters during the past decade and has been well documented (Wu, et. al., 1980; 1982; 1983; 1984). According to past experience, the major attraction of RBC is the ease of operation, low power and maintenance costs, and flexibility. However, the innovative application of RBC for VOC removal is still new.

Since air stripping is a process of providing intimate contact between the contaminant-bearing water and air, a Rotating Disk Contactor (RDC), similarly to a packed tower, can provide this contact. The RDC provides a major gas-liquid contact at the interface between ambient gas and the thin water film on the exposed disc surface. The configuration of an RDC is similar to a 90-degree rotated packed tower. In a packed tower, the water cascades down through the stationary porous media; the rotating discs of a RDC system bring the liquid into the air. There is an important difference between these two systems. The transfer of VOCs in a packed tower occurs in a relatively closed space. The air in the packed tower system becomes saturated with VOCs so readily that large quantities of air have to be blown into the tower to achieve substantial

transfer rates, whereas the RDC is generally applied as an open system. There is a direct transfer of VOCs to the atmospheric boundary, which offers an enormous dilution capacity. Therefore, the driving force of the transfer may maintained at the maximum level.

There are reasons to believe that, under carefully selected operating conditions, the RDC system may be more competitive with countercurrent packed tower air stripping than are other aeration methods. In addition, RDC does not require extra air for stripping, extra energy for liquid transport, and extra support structure, valves, and piping that are normally required in a tall air stripping tower. Precipitation and clogging do not usually cause significant operational problems. The development of biomass and precipitate on disc surface can also be avoided by increasing the disc rotational speed. The ease of replacing media may make the total maintenance cost of the RDC system less than that of a packed tower.

One potential drawback to air stripping is the creation of an emission of volatile organics. The VOCs content in ground, surface or drinking water is usually much less than 1 ppm, therefore, the threat to air quality is minimal. However, the air quality problem becomes severe when treating municipal or industrial wastewaters of high VOC content. When control of these emissions is required by

regulatory agencies, a vapor-phase carbon adsorption process has to be added. In this case, RDC will be covered and auxiliary air will be blown into the system. Emission problem is expected to be most severe for biological treatment processes using aeration devices, such as the activated sludge system. Due to the large open surface area, the control of emission becomes more difficult. The RDC system may be added prior to conventional biological processes as a pretreatment. On the other hand, the RDC has a great potential to be applied as an alternative of other aerobic biological processes. As mentioned before, in a biological treatment system, the interest is in reducing volatilization. In order to minimize the volatilization of VOCs, it is crucial to understand the parameters that control the VOC emission from the RDC.

This study is primarily planned to investigate the technical and economical feasibility of using the RDC technology for VOc removal from the contaminated water by means of physical treatment (air stripping). Therefore, the efforts will be focused on the modeling of the physical VOC transfer in the rotating disk contactor. The following objectives will be studied:

(a). the influence of operational parameters such as rotational speed, depth of immersion, VOC loading, hydrau-

lic loading, and temperature on air stripping efficiency of VOCs in the RDC system,

(b). the development of mathematical models for predicting VOCs transfer in the RDC system,

(c). the cost estimation for RDC capital investment, operation and maintenance in comparison with the conventional packed tower air stripping system.

The relationship between VOC removal and controlling factors will be defined and the criteria employed for future system design will be understood. Although biological effects are not included in this work, the results of this study can still provide an assessment of VOC emission to the atmosphere for decision makers who have to determine the appropriateness of the design and operation of a rotating biological contactor system.

Table 1.1 List of 31 VOCs as priority Pollutants (ES&T, 1979)

Compound	Percent[a] of Samples	Number of[b] industrial categories
Chloroform	40.2	28
Toluene	29.3	28
Benzene	29.1	25
Methylene chloride	34.2	25
1,1,1 Trichloroethane	10.2	25
Ethylbenzene	16.7	24
Trichloroethylene	10.5	21
Tetrachloroethylene	10.2	19
1,2-trans-dichloroethene	7.7	18
1,1 Dichloroethylene	7.7	17
Dichlorobromomethane	4.3	17
1,2 Dichloroethane	6.5	16
Chlorodibromomethane	2.5	15
Carbon tetrachloride	7.7	14
Tetrachloroethane	4.2	13
Bromoform	1.9	12
1,1,2 Trichloroethane	1.9	12
Trichlorofluoromethane	6.8	11
Acrylonitrile	2.7	10
Chlorobenzene	5.0	10
1,1 Dichloroethane	1.4	8
Methyl chloride	1.9	6
Acrolein	1.2	5
1,2 Dichloropropane	2.1	5
1,3 Dichloropropene	1.0	5
Dichlorodifluoromethane	0.3	4
Bis(chloromethyl)ether	0.1	2
Chloroethane	0.4	2
Vinyl chloride	0.2	2
2-Chloroethyl vinyl ether	1.5	1
Methyl bromide	0.1	1

a. a total of 2532 to 2998 samples were analyzed during the EPA's investigation on 31 August 1978.
b. a total of 32 industrial categories.

Table 1.2 Industrial Occurrence of Trichloroethylene (adopted from "Treatability Manual", EPA, 1983)

Industry	Raw wastewater				
	Number of samples	Number of detections	Detected concentrations, µg/L Minimum	Maximum	Mean
Auto and Other Laundries (a)	18	11	<0.5	800	<240
Coal Mining (b)	47	1		3.0	
Inorganic Chemicals Manufacturing (b)	1	1		20	
Leather Tanning and Finishing	18	2	10	20	15
Aluminum Forming	1	1		430	
Battery Manufacturing (h) (j)	53	23	ND	10	<5.7
Coil Coating (k)	43	16	0.0	3,100	170
Electrical/Electronic Components (c)	28	13	6.0	3,500	<300
Foundries	53	13	<10	280	<72
Metal Finishing (b) (h)	124	92	ND	1.3 x 10E5	8,500
Photographic Equipment/Supplies (d)	36	18	0.02	120	14
Porcelain Enameling	1	1		<10	
Nonferrous Metals Manufacturing (f) (h)	95	17	ND	900	59
Ore Mining and Dressing (b)	33	0			
Organic Chemicals and Plastics and Synthetic Resins	33	NA	NA	NA	43
Paint and Ink Formulation (c)	28	19	<5.0	40,000	<2,600
Petroleum Refining (b)	16	0			
Pulp and Paperboard Mills (h)	69	28	ND	850	49
Rubber Processing	1	1		<0.1	
Soap and Detergent Manufacturing (a) (i)	3	3	0.4	27	9.7
Steam Electric Power Plants (e)	11	2	0.57	<4.0	<2.3
Textile Mills (b) (g)	78	24	1.0	5,600	300

NA, not available; ND, not detected. See Section 1.1 Introduction for additional information.

(a) Screening data.
(b) Screening and verification data.
(c) Analytic method not specified.
(d) Screening plus additional data.
(e) Verification data plus surveillance and analysis program data.
(f) Detections >10 µg/L.
(g) Mean calculated using medians.
(h) Minimum, maximum, and mean are based on the number of samples, not detections.
(i) Data presented are for phenanthrene/trichloroethylene.
(j) Detections may include values less than 5 µg/L.
(k) Reference reports 0.0 µg/L for detections less than detection limit 10 µg/L.

Information represents data from the USEPA verification program except as noted.

Table 1.3 Industrial Occurrence of 1,1,1-Trichloroethane (adopted from "Treatability Manual", EPA, 1983)

Industry	Raw wastewater				
	Number of samples	Number of detections	Detected concentrations, µg/L Minimum	Maximum	Mean
Auto and Other Laundries (a)	19	9	<2.0	3,300	<660
Coal Mining (b)	47	4	3.0	23	8.0
Inorganic Chemicals Manufacturing (b)	1	1		0.14	
Iron and Steel Manufacturing (a)	9	7	<10	420	<110
Leather Tanning and Finishing	18	3	<10	10	<10
Aluminum Forming	1	1		530	
Battery Manufacturing (h) (i)	44	31	ND	30	<10
Coil Coating (j)	43	20	0.0	3,100	190
Electrical/Electronic Components (c)	28	12	3.0	7,700	<1,400
Foundries	53	10	<10	16,000	<1,600
Metal Finishing (b) (h)	94	78	ND	1.3 x 10E6	34,000
Photographic Equipment/Supplies (d) (J)	33	13	0.0	1,600	110
Porcelain Enameling (k)	1	1		<10	
Gum and Wood Chemicals	1	1		640	
Nonferrous Metals Manufacturing (f) (h)	58	5	ND	40	3.6
Ore Mining and Dressing (b)	33	9	NA	10	6.7
Organic Chemicals and Plastics and Synthetic Resins	33	NA	NA	NA	100
Paint and Ink Formulation (c)	26	14	5.0	500	<150
Petroleum Refining (b)	16	0			
Pulp and Paperboard Mills (h)	81	35	ND	2,000	130
Rubber Processing	1	1		1.1	
Steam Electric Power Plants (e)	11	1		0.68	
Textile Mills (b) (g)	73	21	2.0	1,200	89

NA, not available; ND, not detected. See Section 1.1 Introduction for additional information.

(a) Screening data.
(b) Screening and verification data.
(c) Analytic method not specified.
(d) Screening plus additional data.
(e) Verification data plus surveillance and analysis program data.
(f) Detections >10 µg/L.
(g) Mean calculated using medians.
(h) Minimum, maximum, and mean are based on the number of samples, not detections.
(i) Detections may include values less than 5 µg/L.
(J) Reference reports 0.0 µg/L for detections less than detection limit 10 µg/L.
(k) 1,1,1-trichloroethane was not detected during the screening program and was not detected in the verification program.

Information represents data from the USEPA verification program except as noted.

Table 1.4 Industrial Occurrence of Ethylbenzen (adopted from "Treatability Manual", EPA, 1983)

Industry	Raw wastewater				
	Number of samples	Number of detections	Detected concentrations, µg/L Minimum	Maximum	Mean
Auto and Other Laundries (a)	17	9	1.5	18,000	<2,500
Coal Mining (b)	48	4	2.0	11	5.0
Iron and Steel Manufacturing (a)	11	8	<3	640	<210
Leather Tanning and Finishing	18	12	<10	150	<78
Aluminum Forming	4	3	15	40	32
Battery Manufacturing (h) (i)	13	2	ND	<10	<6.7
Electrical/Electronic Components (c)	28	11	0.5	110	<20
Foundries	53	8	<10	78	<18
Metal Finishing (b) (h)	87	56	ND	5,500	200
Photographic Equipment/Supplies (d)	24	7	0.74	1,100	220
Gum and Wood Chemicals	3	3	20	37,000	12,000
Pharmaceutical Manufacturing	5	5	10	42,000	8,000
Nonferrous Metals Manufacturing (f) (h)	74	3	ND	21	2.5
Ore Mining and Dressing (b)	32	4	NA	18	6.7
Organic Chemicals and Plastics and Synthetic Resins	56	NA	NA	NA	670
Paint and Ink Formulation (c)	27	19	110	7,800	2,400
Petroleum Refining (b)	16	8	28	810	>170
Pulp and Paperboard Mills (h)	102	27	ND	39,000	690
Rubber Processing	4	4	<0.1	8,600	<2,200
Steam Electric Power Plants (e)	9	0			
Textile Mills (b) (g)	78	47	1.0	19,000	920
Timber Products Processing	5	5	37	2,100	710

NA, not available; ND, not detected. See Section 1.1 Introduction for additional information.

(a) Screening data.
(b) Screening and verification data.
(c) Analytic method not specified.
(d) Screening plus additional data.
(e) Verification data plus surveillance and analysis program data.
(f) Detections >10 µg/L.
(g) Mean calculated using medians.
(h) Minimum, maximum, and mean are based on the number of samples, not detections.
(i) Detections may include values less than 5 µg/L.

Information represents data from the USEPA verification program except as noted. The pollutant was not detected during the screening program and therefore was not sampled in the verification program for the following industries: Coil Coating, Porcelain Enameling.

Table 1.5 Industrial Occurrence of Tetrachloroethylene (adopted from "Treatability Manual", EPA, 1983)

Industry	Raw wastewater				
	Number of samples	Number of detections	Detected concentrations, µg/L Minimum	Maximum	Mean
Auto and Other Laundries (a)	27	20	0.5	93,000	<11.000
Coal Mining (b)	47	0			
Inorganic Chemicals Manufacturing (b)	1	1		36	
Iron and Steel Manufacturing (a)	9	5	<10	1,200	<370
Leather Tanning and Finishing	18	4	10	150	56
Aluminum Forming	4	4	5.0	<4,000	<19,000
Battery Manufacturing (h) (i)	18	5	ND	<10	<6.0
Electrical/Electronic Components (c)	28	15	0.2	800	<120
Foundries	53	9	<10	370	<110
Metal Finishing (b) (h)	115	78	ND	1.1 x 10E5	4,200
Photographic Equipment/Supplies (d)	21	10	0.01	96	8.9
Porcelain Enameling	4	2	<10	<10	<10
Nonferrous Metals Manufacturing (f) (h)	95	18	ND	310	15
Ore Mining and Dressing (b)	33	2	NA	11	7.8
Organic Chemicals and Plastics and Synthetic Resins	19	NA	NA	NA	5,100
Paint and Ink Formulation (c)	29	21	<5.0	4,900	<680
Petroleum Refining (b)	16	1		>50	
Pulp and Paperboard Mills (h)	96	17	ND	220	13
Rubber Processing	1	1		1.4	
Soap and Detergent Manufacturing (a)	1	1		15	
Steam Electric Power Plants (e)	11	2	0.4	<10	<5.2
Textile Mills (b) (g)	78	24	1.0	1,100	180

NA, not available; ND, not detected. See Section 1.1 Introduction for additional information.

(a) Screening data.
(b) Screening and verification data.
(c) Analytic method not specified.
(d) Screening plus additional data.
(e) Verification data plus surveillance and analysis program data.
(f) Detections >10 µg/L.
(g) Mean calculated using medians.
(h) Minimum, maximum, and mean are based on the number of samples, not detections.
(i) Detections may include values less than 5 µg/L.
(j) Tetrachloroethylene was not detected during the screening program, and was not detected in the verification program.

Information represents data from the USEPA verification program except as noted. The pollutant was not detected during the screening program and therefore was not sampled in the verification program for the following industries: Coil Coating.

Table 1.6 Industrial Occurrence of Chloroform (adopted from "Treatability Manual", EPA, 1983)

Industry	Raw wastewater				
	Number of samples	Number of detections	Detected concentrations, µg/L Minimum	Maximum	Mean
Auto and Other Laundries (a)	28	23	0.7	35,000	<2,700
Coal Mining (b)	47	25	3.0	480	95
Inorganic Chemicals Manufacturing (b)	1	1		85	
Iron and Steel Manufacturing (a)	34	29	<3	1,400	<110
Leather Tanning and Finishing	18	12	2.0	41	<17
Aluminum Forming	34	32	2.0	12,000	<400
Battery Manufacturing (h) (i)	13	6	ND	<10	<10
Electrical/Electronic Components (c)	28	15	5.5	50	<15
Foundries	53	11	1.0	470	<79
Metal Finishing (b) (h)	101	84	ND	690	31
Photographic Equipment/Supplies (d)	22	11	1.0	26	10
Porcelain Enameling	4	4	<10	<10	<10
Explosives Manufacturing	1	1		540	
Gum and Wood Chemicals	3	3	10	1,100	530
Pharmaceutical Manufacturing	6	6	10	2.8 x 10E5,	35,000
Nonferrous Metals Manufacturing (f) (h)	95	32	ND	1,800	48
Ore Mining and Dressing (b)	32	9	NA	35	7.6
Organic Chemicals and Plastics and Synthetic Resins	58	NA	NA	NA	240
Paint and Ink Formulation (c)	29	17	<5.0	900	<150
Petroleum Refining (b)	16	9	<5.0	100	<41
Pulp and Paperboard Mills (h)	154	103	ND	9,700	510
Rubber Processing	5	5	1.9	27	8.5
Soap and Detergent Manufacturing (a)	2	2	1.1	4.8	3.0
Steam Electric Power Plants (e)	11	3	0.17	<10	<3.9
Textile Mills (b) (g)	78	34	1.0	640	77
Timber Products Processing	5	5	10	20	12

NA, not available; ND, not detected. See Section 1.1 Introduction for additional information.

(a) Screening data.
(b) Screening and verification data.
(c) Analytic method not specified.
(d) Screening plus additional data.
(e) Verification data plus surveillance and analysis program data.
(f) Detections >10 µg/L.
(g) Mean calculated using medians.
(h) Minimum, maximum, and mean are based on the number of samples, not detections.
(i) Detections may include values less than 5 µg/L.

Information represents data from the USEPA verification program except as noted. The pollutant was not detected during the screening program and therefore was not sampled in the verification program for the following industries: Coil Coating.

Table 1.7 Occurrence of VOCs in POTW Influent (EPA, 1982)

Compounds	Influent			Effluent		
	Frequency detected	Concentration (ug/l)		Frequency detected	Concentration (ug/l)	
	%	Min.	Max.	%	Min.	Max.
TCE	90	1	1,800	45	1	230
PCE	95	1	5,700	79	1	1,200
CLF	91	1	430	82	1	87
EB	80	1	730	24	1	49
TCEN	85	1	30,000	52	1	3,500

Table 1.8 Occurrence of Selected VOCs in Surface Water (EPA, 1981)

	No. Sites Sampled	Positive Samples %	Mean ug/l	Max. ug/l
Trichloroethylene	133	32.2	0.47	3.2
Tetrachloroethylene	180	12.8	1.49	21.0
Chloroform	144	35.7	3.46	30.0
1,1,1-Trichloroethane	133	16.5	0.56	3.3
Ethylbenzene	NI	NI	NI	NI

Table1.9 Summary of VOC Occurrence Data (EPA, 1981)

	No. States	No. Wells Tested	positive Wells %	Max. Conc. ug/l
TCE	8	2,894	28	35,000
PCE	5	1,652	14	50
TCEN	3	1,611	23	401,300
CLF	1	372	11	67
EB		NI	NI	NI

NI: not investigated

Table 1.10 Selected VOCs in Raw and Finished Ground Water (EPA, 1981)

	Positive Samples %		Concentration (ug/l) Mean		Range	
Chemical	Raw	Fin	Raw	Fin	Raw	Fin
TCE	38.5	36.0	29.72	6.76	0.2-125.0	0.1-53.0
PCE	18.5	22.0	0.98	2.08	0.1- 2.0	0.2- 3.1
TCEN	23.1	21.7	4.8	2.13	0.3- 13.0	1.3- 3.0
CLF	NI	NI	NI	NI	NI	NI
EB	NI	NI	NI	NI	NI	NI

Table1.11 Selected VOCs Found in Drinking Water Well (CEQ, 1981)

Chemical	Concentration ug/l	State	Highest surface waterconc. reported ug/l
TCE	27,300	PA	
	14,000	PA	
	3,800	NY	160
	3,200	PA	
	1,530	NJ	
	900	MA	
PCE	1,500	NJ	
	740	CT	21
	717	NY	
CLF	490	NY	
	420	NJ	700
	67	NY	
TCEN	5,440	ME	
	5,100	NY	5.1
	1,600	CT	
	965	NJ	
EB	2,000	NJ	NI

CHEPATER 2 LITERATURE REVIEW

Previous studies on treatment techniques indicated that adsorption processes using activated carbon (Parmele, 1986; Lucey, 1984, Weber, 1984) or adsorbent resins, chemical oxidation with ozons, or chlorine dioxide, UV/hydrogen peroxide, reverse osmosis (Sorg, 1984), and steam stripping (IT Co., 1984) are capable of removing 99% VOCs from the contaminated source of potable water supplies or groundwater but these treatment alternatives are very expensive and operationally complex (Lucey, 1984; Love, 1984). A summary of the removability and treatability studies of the VOCs were US collected in EPA's "Treatability Manual" (US EPA, 1984). Based on that manual, biological treatment processes and air stripping seem to be the two most promising alternatives (Weber, 1984, Rittmann, 1980, Clark, 1984).

2.1 Biodegradation of VOCs

Biological processes are, in general, the most cost effective techniques for treating aqueous wastewater containing organic constituents. They have been proved to be capable of treating a wide variety of industrial wastes with limited environmental impact. The biodegradation studies associated with VOCs have currently proposed the most controversial results. A number of VOCs, such as

chlorinated benzenes, methylene chloride, and toluene are, indeed, removed through biodegradation (Namkung, 1987, Callahan, 1979). However, most studies showed that biological treatment of low molecular weight VOCs, such as trichloroethylene, tetrachloroethylene, trichloroethane, and chloroform, etc., is difficult in either aerobic or anaerobic environments and, therefore, the treatment efficiency is low (National Research Council, 1984; Nelson, 1986; Wilson, 1985).

Field studies using reclaimed wastewater injected into an aquifer indicated that TCE exhibited a half-life transformation as long as 300 days (Robert, 1980). Barrio-Lage (1987) studied the depletion of VOCs in water and natural sediments and reported the first-order rate constants as 3.2 to 8.7x 10^{-4} for TCE. Bouwer (1982) studied the adsorption and biodegradation by activated carbon and fixed-film bacteria and pointed out that CLF, 1,1,1 TCEN and PCE were weakly adsorbed and not biodegraded aerobically.

Biodegradation of halogenated aliphatic hydrocarbons was studied by Bouwer, Rittmann, and McCarty (1981) in both aerobic and anaerobic environments. Their study stated that the low molecular VOC such as TCE at the concentration commonly found in the surface and groundwater was not

degraded aerobically but slightly degraded under oxygen-free conditions. Carbon-14 tracer technique indicated that the initial steps in the degradation of those halogenated aliphatic hydrocarbons appeared to be reductive dehalogenation. The experimental data indicated that anaerobic degradation may occur as:

Carbon tetrachloride --► Chloroform --► Methylene chloride
Tetrachloroethylene---► Trichloroethylene---► Dichloroethylene --► Vinyl chloride
1,1,1-Trichloroethane---► 1,1-dichloroethane --► Chloroethane

Additional work in degradation was conducted by Nelson et. al.(1986) Their work showed that aerobic metabolism of TCE by a pure culture (designated strain G4) was obtainable. Physiochemical and morphological studies of strain G4 was ascribed to the genus Acintobacter. The same study also illustrated that no growth of strain G4 was found when growth substrates including glucose, acetate, lactate, succinate and ethanol were utilized. This means that the process of co-metabolism between TCE and the organic co-substrates mentioned above does not occur in the G4 culture. Wilson et. al. (1985) investigated biotransformation of TCE in sandy soil packed column and successfully demonstrated the conversion of TCE to carbon dioxide aerobically if the soil was exposed to natural gas to

enrich for microorganisms that oxidize ethylene and the other small alkanes present in natural gas. These studies additionally showed that even at the micronutrient level, the removal of VOC by biological treatment always required a considerably long reaction time or a relatively large bio-reactor.

The removal mechanisms of VOCs in the activated sludge system were studied extensively by Weber (1984), Blackburn (1985), Kincannon (1983), Petrasek (1983) and Namkung (1987) etc. Petrasek (1983) studied the removal of 16 VOCs in POTW and found that most of the VOCs investigated had removal greater than 95%. Only 1,1,2 trichloroethane and dibromochloromethane had mean removals less than 90 percent. According to his data, average removals of 24 percent were obtained in the primary clarifier. Because of the low concentrations found in primary sludge, he concluded such removals were due to desorption, and partitioning to the solids was not a significant removal mechanism. The stripping of VOCs from the aeration basin was the principal removal mechanism. However, sorption to the biomass was found to be sufficiently rapid to influence the fraction of VOCs being stripped. Due to the low VOCs' concentrations observed in the waste activated sludge, the adsorbed compounds would have to have been biodegraded.

The results obtained from POTWs by Namkung (1987)

indicated that 94% to 95% of chlorobenzene, ethylbenzene, benzene, methylene chloride, and toluene would be removed by biodegradation in a conventional activated sludge treatment. However, chloroform, 1,2 dichloroethane, tetrachloroethylene, trichloroethylene, and 1,1,1 trichloroethane have zero or very low biodegradation rates. The volatilization became the main removal mechanism. Adsorption was a negligible removal mechanism for both biodegradable and nonbiodegradable VOCs. An estimation of VOC removals by these three mechanisms was adopted from Namkung's study and listed in Table 2.1. Kincannon (1983) and Blackburn (1985) also reported similar results from lab-scale activated sludge processes. Because relatively higher air/water ratios were used in these studies, the experimental data showed complete removals of tetrachloroethylene, 1,2-dichloroethane, and 1,1,1-trichloeoethane by volatilization. The biodegradability of the VOCs will influence the extent to which the compound is removed by air stripping. The more biorefractory the compound, the more likely the compound is to be stripped.

Abrams et al. rated trichloroethylene and tetrachloroethylene as very difficult to biodegrade (Abrams, 1975). However, by using the static-culture, flask-screening techniques, Tabak (1981) reported that significant degradation with gradual adaptation occurred for trichloroethylene and tetrachloroethylene, and slow biodegradative activity

occurred for 1,1,1-trichloroethane. He pointed out that the concentrations used were significantly higher than those may frequently occur in normal wastewaters. The success of biological treatment processes requires a sufficiently large population of acclimated microorganisms to degrade the pollutants and a supply of substrates as an energy source to sustain the population. The low concentrations of VOCs usually found may retard an induced enzyme formation by the microoganism and the formation of an acclimated microbial population to degrade the contaminants. The high volatilization losses of VOCs may further decrease their concentrations to hinder the biodegradation from occurring. Therefore, an ideal biological process should be capable of maximizing the amount of biomass in the system and minimizing the volatilization losses of VOCs. Bouwer (1985), Stratton (1983) and Namkung (1983) suggested a secondary utilization mechanism that enabled rapid biodegradation of individual trace-level compounds. By definition, a secondary substrate is also an electron donor, but, because of its low concentration, it can not provide enough energy for bacterial growth and maintenance. A primary substrate is necessary to sustain a steady state biomass to allow secondary utilization to occur. It is doubtful if a groundwater or drinking water can supply the required primary substrate. The applications of biological processes for VOCs removal may have to be limited to treat

wastewater.

2.2 Air Stripping

Previous studies have suggested that the transfer of VOCs to the atmosphere by air stripping may be a convenient and potentially cost-effective method to remove these compounds from the low contaminated water, e.g. drinking water, surface water, ground water (Smith, 1980; Symons, 1981, Singley, 1983, Roberts, 1983). Several aeration methods have been applied for this purpose: mechanical surface aeration, diffused aeration, spray or tray towers, spray fountains, and countercurrent packed towers. Among them, the countercurrent packed tower has recently received the most considerable attention. Although air stripping is difficult to optimize for waters containing a spectrum of volatile and nonvolatile components, the process has been most frequently selected as an alternative for the removal of VOCs from contaminated ground water or for the control of THM in drinking water supplies. In addition, it does have appeal as pretreatment prior to another process, such as adsorption, to extend the life of the sorbent by removing sorbable VOCs (O'Brian, 1984; McKinnon, 1984; Amy, 1987).

Some other considerations may include: the efficiency of application in colder climates, whether excessive

evaporative water losses would occur in water-short areas (for water supply), impact of secondary air pollution by emitted VOCs, whether uncontrollable increases in the corrosivity of the water might occur, i.e. impact of dissolved oxygen concentration (Culp, 1984). Among them, the main drawback is that the VOCs are transferred from the water to the atmosphere. When control of these emissions is required, a vapor-phase carbon adsorption process is usually added to the air stripping unit. Parmele (1986) suggested that the expensive add-on adsorption unit often made the total cost two to three times higher than the air stripping unit alone. However, Byers (1988) claimed that for long term operation, e.g. 2 yrs and more, air stripping coupled with an emission control unit would still result in the lower total cost.

2.2.1 Diffused Aeration and Mechanical Surface Aeration

Air stripping of VOCs by diffused aeration has been investigated in pilot scale as well as full scale applications. The stripping efficiencies vary with air-to-water ratio (A/W), hydraulic retention time, water depth, type and arrangement of the diffusers. The pilot-scale and field-aeration studies conducted by US EPA Drinking water Research Division showed 25 to 100% TCE removal efficiencies in a 1.5-in diameter, 4 ft long glass column with 10

min retention time and air-to-water ratios ranging from 1:1 to 20:1 (Nebolsine, 1978). Using a 30-in column, 10 ft in length with five diffusers, the removal efficiency of TCE varied from 69 to 90% with A/W ranged from 69 to 90% (Nebolsine, 1980).

A series of diffused aeration tests conducted at Peace River Water Treatment Plant demonstrated that the removal efficiencies of THMs increased with increasing A/W ratio and water depth (ES&E, 1982). A performance test of the Kenics aerator showed that the effect of increasing A/W ratio diminished as the air flow increased. Increasing the A/W ratio beyond 30:1 would not significantly improve the performance. The removal of THMs had a linear relation with retention time. However, the slope of the correlation line was shallow. The increasing detention time would have limited enhancement of the efficiency (ES&E, 1981). The stripping efficiency in diffused aeration can also be improved by producing smaller bubbles, improving contact basin geometry, or using turbines to reduce bubble size and increase bubble-water contact time. Roberts (1982) proposed a model to predict the stripping efficiency of bubble aeration in a completely mixed, clean water tank,

$$\frac{C_i}{C_e} = 1 + \frac{H_c G}{L}\left(1 - \exp\left(-\frac{K_L a V}{H_c G}\right)\right)$$

Ci and Ce are the influent and effluent concentrations, Hc

is the Henry's constant, G, L are the volumetric gas and water flow rates, V is the water volume. K_La is the overall mass transfer rate constant.

Bilello (1986) suggested that air stripping by diffused aeration was less effective than by packed tower. Diffused aeration improved with increased A/W and water depth. From a practical viewpoint, the cost of operating compressors increases as the A/W ratio is increased. Therefore, increasing the water depth is preferred.

Aeration and air stripping by mechanical surface aerator were studied by Boon (1979) and Roberts (1983) in full-scale and pilot-scale units. The results of their studies shows a direct proportionality between the mass transfer coefficient and power input, as

$$K_La = K (P/V)$$

where P = power input, V = aeration basin volume, and K = proportionality constant. This relation has been found to be valid for mechanical surface aeration systems of widely different sizes, i.e. 0.01 to 1,000 m^3 basin volume (Roberts, 1982; 1983a, 1983b). The proportionality constants are found to be 2.9×10^{-5} for oxygen transfer and about 1.6×10^{-5} for VOCs. The overall water-air mass transfer in the system was based on the concept of staged, completely mixed flow reactor (CMFR) (Boberts, 1983a). The

effluent solute concentration was equated as

$$\frac{C_i}{C_e} = 1 + K\frac{R_L}{R_L + R_G}\frac{P}{L}$$

where R_L, R_G = liquid side and gas side resistances, L = water flow rate, Ci, Ce = influent and effluent concentrations. A comparison between packed tower aeration and surface aeration, shows that surface aeration entailed a greater land area than the packed tower. However, if land is available at low cost, the capital investment is likely to be lower for mechanical surface aeration. Although, power expenditure is a little higher for the surface aeration, the lower operating and maintenance labors cost will make the total operation cost more competitive to packed tower. Surface aerator will be compared more favorably especially where there is a risk of chemical precipitation and biological growth that would tend to clog a packed tower, necessitating shutdown for cleaning (Roberts. 1983a).

2.2.2 Packed Tower

Previous studies have indicated that packed tower air stripping is more economical for removing VOCs than diffused aeration or carbon adsorption(Gross, 1985; Hess, 1981; Symons, 1981). A literature review of current packed tower air stripping practices indicated that removal efficiencies of 90 to 99.99 percent can be attained easily

in pilot tower experiment and in full scale application (Hess, 1981; Hands, 1986; Bilello, 1986; Amy, 1986). A basic design equation can be written as:

$$Z = (HTU)(NTU)$$

$$HTU = U/K_La$$

$$NTU = \frac{R1}{R1-1} \ln \frac{(Xi/Xe)\ (R1-1) + 1}{R1}$$

where Z = packing height, HTU = height of the transfer unit, NTU = no. of transfer units, U = superficial velocity, R1 = stripping factor, Xi, Xe = influent and effluent solutes mole fraction. Many empirical or theoretical equations or models have been proposed to predict the value of K_La (Ponter, 1986). Sherwood and Holloway (1940), Shulman (1955), and Onda (1968) are the models most widely used to date in studies of volatile organics stripping for water treatment. They correlated the mass transfer coefficient with dimensionless groups, such as Schmidt, Sherwood, Reynolds, Froud, and Weber numbers.

The key parameters which directly determine the efficiency of the packed tower system are: volumetric ratio of air to water (A/W), volume of packing required (surface loading rate), effective interfacial area (type of packing medium to be used), and height and diameter of the tower. The air to water ratio defines the volume of air required to remove the desired amount of the VOC contaminants. The

net mass transfer rate increases with increasing air/water ratio until an optimum ratio is reached. However, increasing of the ratio will also increase the pressure drop. Under normal conditions, the tower height and pump power decrease with increasing air/water ratio; the tower diameter and blower power increase with increasing air/water ratio. The optimum ratio of air to water reported by the EPA laboratory is 8: 1 in order to remove an influent TCE concentration of 338 ug/l to an effluent value of 4 ug/l (Love, 1982). In practical application, the air/water ratio is usually selected in the range of 15 to 30 (Hand,1986).

The interfacial area is a complex function of the shape, size, surface characteristics of the packing and air as well as water flow rate. In large scale apparatus the commonly used packings include Raschig rings, Berl saddles, and Pall rings. The materials of the packing include stoneware, porcelain, carbon, plastic, and metal. The sizes of the packing range from 6 mm to 90 mm. Although some authors have correlated the effective interfacial area with superficial liquid velocity, Reynolds number and other hydraulic characteristics, this area is difficult to measure and is usually unknown (Hikita, 1960; Bravo, 1982). For the purpose of liquid/gas contact, a higher media surface area provide a thinner liquid film and hence

enhances contact. But the higher packed media surface area also increases the pressure drop through the column that will influence the size of the packed tower and the ratio of air to water to be used. Sufficient tower height must be provided to ensure that the liquid and gas reach equilibrium. Therefore, the height of the tower is based on the amount of driving force that is available to drive the operating relationship toward equilibrium.

Cost of air stripping has been estimated by mathematical design models or laboratory and full-scale pilot plant studies (Clark, 1984; Hand, 1986; Zaghloul, 1987). When treating groundwater by packed tower, based on a design life of 20 years, the capital cost accounts for 41.7 percent of the total cost, the operational cost accounts for 48.6 percent, and the maintenance cost accounts for 9.7 percent of the total cost (Hand, 1986). Due to the height of the tower, the support structure, valves, and piping accounts for the largest portion, 35 percent, of the capital cost. The pump power requirements for pumping the water to the top of the tower accounts for more than 63 percent of the operational cost (Hand, 1986).

Most of the packed towers experienced problems associated with precipitation of inorganics onto the packing media when treating groundwater. The precipitate causes operational problems like clogging, increasing

pressure drop etc. Although the general maintenance cost of the packed tower is relatively low, the precipitate makes the expensive heavy maintenance inevitable. So far, the application of packed tower stripping to industrial wastewater has not been evaluated, and the impact of microorganism slime growth on the surface of the packing has not been investigated either. It is expected that the precipitate and biomass will cause much more severe problems in treating wastewater than in treating groundwater. In such cases, other aeration methods have to be considered.

Diffused aeration and countercurrent packed column represent the two extremes in the mass transfer processes. Compared to the packed tower, diffused aeration provides less interfacial area for mass transfer but greater gas-liquid contact time. The packed tower provides a greater effective area but lower liquid volume. Literature indicates that diffused aeration provides an optimum treatment system for the absorbance of gas in the water, packed tower provides an optimum system for the removal of VOCs.

2.3 Rotating Disk Reactor (RDC)

A Rotating Disk Contactor consists of a series of

thick, vertical, large-diameter circular disks mounted on a rotating horizontal shaft, set atop a concrete trough or semi-circular steel tank. The dimensions of a full scale commercial RDC may be seen in Table 2.2. The disks are made of polystyrene, polyethylene, polyvinyl chloride and fiberglass (Strom, 1985). In order to increase the surface area, the disks surface may be corrugated, honeycombed, or otherwise formed instead of flat. In most applications in biological treatment the disks, positioned parallel or perpendicular to the flow, rotate slowly with approximately 40% of the disk surface area submerged in the wastewater. The rotation alternatively exposes the disk surface to the atmosphere above the tank, and subsequently resubmerges it into the wastewater, thus, achieving mass transfer between bulk wastewater, atmosphere, and biofilm. A microbial film about 2-3 mm thick develops on the disk surface. Excessively grown biomass is sloughed off by the shearing force as the disks move through the wastewater. The rotation also serves to keep the tank contents mixed, disperse the excess sloughed biomass, and increase the surface turbulence of the bulk water.

The general advantages of the RDC system in biological treatment application includes: great flexibility, ease of operation, low operating and maintenance costs (Sullivan, 1984); smaller reactor size, high resistance to shock and hydraulic loadings, low sludge production with good sludge

settleability and dewaterability (compared with activated sludge), low head loss, no clogging or ponding problem (compared with trickling filters), less foaming, aerosols, and air stripping (compared with activated sludge). Due to these advantages, RBC has extreme attraction to small scale applications, developing countries or rural areas. The major disadvantages may be the lack of operational control. However, this characteristic contributes to its major advantage-the simplicity of operation. Other disadvantages and problems include: lack of manufacturing and operating experience, disc, shaft, and drive failure due to mechanical deficit or overloading (Poon, 1984), brittleness of media exposed to sunlight, media becoming loose or cracking, difficulty and complexity of kinetic evaluation, modeling, scale-up, and optimization. Most of these disadvantages or problems actually come from the fact that RDC is a relatively new technique with only brief service history and experience (Antonie, 1974; Clark, 1978; Wu, 1981; Strom, 1985). Most of these early operational problems are now corrected (Poon, 1984; Jaffer, 1984; Brenner, 1984).

The main controlling factors which greatly influence the RDC performance as a biological treatment process include influent wastewater characteristics, hydraulic retention time (Zogorski,1984), surface organic loading

(Kincannon, 1984; Stover, 1982), disc rotational speed, depth of submergence and temperature (Forgie, 1984),and some of these factors are also important to the air stripping operation. Current RBC design for biological treatment of municipal and industrial wastewaters normally maintains the disc rotational speed at 1-2 rpm and the submerged discs' area equal to 40% of the total surface area of the fixed film media. This means that for a given wastewater temperature and composition, the RBC performance is mainly affected by the rate of disc surface organic loading (Wu, 1980).

Since RDC has been traditionally employed as a biological treatment alternative, the mathematical models are mainly developed for predicting the removal of carbonaceous and nitrogenous substrates. The complexity of the interaction between the biomass and solid, liquid, and gaseous phases makes the kinetic evaluation and modeling, scale-up and optimization very difficult. The existing models mostly include empirical models based on multiple regression techniques (Wu, 1982a, 1982b) or kinetic models based on apparent reaction and growth kinetics (Harremois, 1982; Peters, 1984).

More recently, some authors aware of the importance of both liquid film and biofilm in describing fixed film removal kinetics, have incorporated mass transfer

principles into RBC models. Famularo (1978), Mueller (1980a, 1980b) considered the material balance in the tank, biofilm and liquid film of the RBC system, combining advection transport, diffusion transfer, and biological kinetics into their models (Peters, 1984a). Atasi and Borchardt (1982) used the dimensionless Damkohlor number, to determine if the overall substrate removal is mass transfer limited or reaction limited.

The general approaches usually consider several steps involved in mass transfer between biofilm, air and bulk water. In the submerged disk fraction, these steps have been considered:

1. Transport of the substrates from the bulk water to the surface of the biofilm by convection and diffusion.

2. Internal diffusional transfer of the substrates through the biological film.

3. Biological utilization of the soluble substrate by the biomass.

4. Diffusion of part of the reaction products to the bulk of the water.

When the submerged fraction is brought up to air by the disk rotation, a thin water film will be formed on the disk

surface. The mass transfer steps involved in the overall absorption and desorption processes are:

1. Components transport in the liquid film adhering onto the disk surface and transfer between the air and liquid film interface.

2. Components transfer at the interface between liquid film and biofilm.

3. Components transfer in the biofilm.

4. Components transfer between the bulk liquid and submerged disks and transport in the bulk water.

It is believed that even at low Reynolds numbers, i.e. low rotational speeds in the RDC system, the convection always predominates in the bulk water (Shieh, 1981). A uniform concentration of each component in bulk water is then expected. It is also assumed that the biofilm or water film is stagnant on the disk surface with uniform thickness to simplify the model development.

In the studies of physical mass transfer in the RDC system, the biochemical reactions and biofilm diffusion are excluded. Since the innovative application of RDC for VOC removal is still new, not much information can be found

in the literature. Previous works on physical air-water mass transfer were concentrated on the absorption of oxygen in the RBC system. Yamane and Yoshida (1972) considered diffusion transfer on the exposed disks as the dominant oxygen absorption mechanism, and solved the Fick's law

$$\frac{dC}{dt} = D\frac{\partial^2 C}{\partial x^2}$$

With the initial and boundary conditions,

$$\begin{array}{lll} t = 0, & 0 < x < \delta, & C = C_i, \\ t > 0, & x = \delta, & C = C_s \\ t > 0, & x = 0, & \dfrac{\partial C}{\partial x} = 0 \end{array}$$

the liquid phase oxygen transfer coefficient was equated as

$$\frac{C_a - C_i}{C_s - C_i} = \sum_{n=0}^{\infty}(-1)^n \mathrm{erfc}\left\{\frac{(2n+1)\delta - x}{2\sqrt{Dt}}\right\} + \sum_{n=0}^{\infty}(-1)^n \mathrm{erfc}\left\{\frac{(2n+1)\delta + x}{2\sqrt{Dt}}\right\}$$

when $\dfrac{\delta}{\sqrt{Dt_a}} > 1.7$, and $\dfrac{\bar{k}_L}{2\sqrt{D/\pi t_a}} = 1$

where D = diffusivity, ta = exposure time, δ = thickness of the water film on the disks.

Chesner and Molof (1976, 1977) also suggested that rotational speed was the better scale-up factor than peripheral velocity, the previously used design parameter. They found that with the same peripheral velocity, smaller disk diameter with higher rotational speed had better oxygen transfer and organic removal efficiencies than the

larger diameter with lower rotational speed.

Bintanja et al. (1975) employed the Yamane & Toshida's solution and concluded

$$K_L = 2\sqrt{\frac{D}{\pi t_a}} \qquad \text{when} \frac{\delta}{\sqrt{Dt_a}} \geq 1.7$$

$$K_L = \frac{\delta}{t_a} \qquad \text{when} \frac{\delta}{\sqrt{Dt_a}} < 0.8$$

According to their data, experimentally determined from a flat-disk RDC pilot plant, the experimental K_L values were 49% to 87% of the theoretical values. Zevalkink (1979) attributed these deviations to the incomplete mixing between the water film and the bulk water in the trough. He extended the work of Bintanja (1975) and derived the following:

$$0.8 < \frac{\delta}{\sqrt{Dt_a}} < 1.7$$

$$\bar{k}_L = 2\sqrt{D/\pi t_a}\,(1 - 4.21\exp(-3.2t_a))$$

Zevalkink (1978) solved the Navier-Stokes equation by analogizing flat plate withdrawal and determining the liquid film thickness adhering on RDC disks as:

$$\phi = \arccos(R_i/R),$$

$$\delta = \frac{4}{15}\sqrt{\frac{2\nu}{g}}\sqrt{wR\sin\phi}$$

For water at 20 C, the water film thickness can be expressed as

$$\delta = 1.2 \times 10^{-4}\sqrt{V_c}$$

Famularo (1978) applied the film thickness, derived by Levich (1962), for the vertically withdrawn flat plates, and proposed

$$\delta = 6.85\, v^{2/3} + \alpha$$

$$K_L = 2\frac{D}{\delta}$$

Friedman et. al (1979) confirmed the significance of rotational speed and presented an empirical equation for oxygen transfer

$$\ln K_L = 1.31 \ln w + 14.78$$

where the units of K_L and w were 10^{-6} m/s and rpm respectively.

Kim and Molof (1980) correlated the oxygen transfer coefficient with the volume renewal number, N_V.

$$K_L a = a\ (N_V)^b$$

$$N_V = w^{1.5}\ R^{0.5} / S$$

They concluded that the oxygen transfer coefficient,

K_La was proportional to $D^{0.4}$ when rotational speed was held constant.

K_La was proportional to $w^{1.1}$ when the disk size was held constant.

K_La was proportional to $S^{0.9}$.

where D = disk diameter, w = rotational speed, S = half-space between disks. The semi-empirical equation for oxygen transfer was

$$K_La = 0.0011\ (N_V)^{0.732}$$

Matsuo and Yamamoto (1985) incorporated an interchange coefficient concept into the basic models of Yamane and Yoshida (1972), Bintanja (1975), and Zeevalkink (1979). They introduced the interchange coefficient, f, as an index to represent the extent of mass exchange between the water film and bulk liquid. Thus, the apparent K_L, K_L^*, would be a fraction of the theoretical K_L

$$K_L^* = \frac{f}{(1-f)k + f} K_L$$

$$f = f1 + (1-f1)\ f2$$

$$f_1 = \frac{\sqrt{Dt_w}}{\delta}\left(\frac{1}{\sqrt{Dt_w}} - \text{ierfc}\left(\frac{\delta'}{\sqrt{Dt_w}}\right)\right)$$

$$f_2 = a_t\, w^{1.3} R^{2.1} (\sin(\phi))^{1.5}$$

where tw = retention of the disk submerged in bulk water, δ' = thickness of the disk surface boundary film in bulk water, a_t = constant.

Severin (1980) correlated the oxygen mass transfer coefficient with the Reynolds number calculated from tank geometry:

$$K_L a = \frac{K1\ K2}{m\ Cs}\, \mu\ (n'\ (R^2 - R_i^2))^{0.815} \left(\frac{w\ f_v}{\nu\ p\ q}\right)^{1.815}$$

$$N = 5.7 \times 10^{-8}\ Re^{1.815}$$

where K1 = experimental constant, 5.7×10^{-8}, K2 = conversion factor, 35.32 mg ft^3/ g L, f_v = void volume, p = wetted perimeter, q = number of disk per length, m = area of disk per area of circle with the same radius, n' = number of disk, R = radius of the disks, R_i = distance between shaft to water surface, Re = Reynolds number. He pointed out that scale up by tip speed and surface area overestimated the transfer capacity of full-scale plants, and by rotational speed underestimated the capacity of large diameter plants.

Ouano (1978, 1981) considered the agitated surface of the bulk water as the major mass transfer interface and

derived the equation by dimensionless analysis procedure.

$$\frac{K_L\ V/At}{D} = K'\left(\frac{At}{Ap}\right)^{a'}\left(\frac{R^2\ w\ p}{\mu}\right)^{b'}$$

where At = surface area of the bulk water, Ap = projected area of media on water surface, K', a', b' = empirical constants.

Table 2.1 Estimates of VOC removals in Activated Sludge [a]

Compound	percent removal by each mechanism Volatil.	Adsorpt.	Biodegrad.
Benzene	2.6	0.0	94.8
Chlorobenzene	2.3	0.1	95.1
Ethylbenzene	3.2	0.1	94.1
Methylene Chloride	1.5	0.0	96.0
Toluene	3.2	0.1	94.2
Chloroform	43.6	0.2	-
1,2-Dichloroethane	20.0	0.1	-
Tetrachloroethylene	83.6	0.4	-
1,1,1-Trichloroethane	44.9	0.3	-
Trichloroethylene	69.4	0.2	-

a. Namkung, E., JWPCF, 59, 7, 1987

Table 2.2 Dimensions of Full Scale RBC Unit [a]

width of trough	3.81 m (12'-6")
length of trough	8.03 m (26'-4")
diameter of discs	3.66 m (12')
depth of trough	1.70 m (5'-7")
total disc surface area	16,722 m^2 (180,000 sq.ft)
distance between discs and trough	0.08 m (3")

a. standard high media density unit (Lyco, Inc.)

CHAPTER 3 THEORIES

In a rotating disc contactor, the air-water contact is established by means of discs vertically rotating on a horizontal central shaft. The scheme of RDC is illustrated as Figure 3-1. As a consequence of the rotation, water from the trough is brought upwards into the air continuously, and spread as thin films on the surface of the discs. Oxygen and VOCs can transfer between the water and atmosphere in two ways. First, during the transport of the the water film on the discs, oxygen will diffuse through the air-water boundary into the water film. In a similar way, VOCs will diffuse into the surrounding air. Second, they can also transfer through the agitated water surface of the bulk water in the basin. We consider physical absorption/desorption in which no chemical reaction takes place among the dissolved oxygen, VOCs and the water. For a continuous flow RDC system, concentrations of VOCs throughout the RBC system can be obtained by solving material balance equations for perfectly mixed regions.

Accumulation Rate within the System	=	Rate of VOC Flow into the System	-	Rate of VOC Flow out of the System	+	VOC Sink/Source Rate

$$V\frac{dC}{dt} = QC_0 - QC + VR_s \tag{3.1}$$

where Rs is the rate of mass transfer per unit volume of

liquid; Q is the water flow rate; and Co and C are the influent and effluent concentrations respectively.

For transfer from the liquid (water) to the gas (air), the steady state mass transfer rate per unit volume of liquid is usually defined by the relation

Rate of Mass Transfer per Unit Volume of Liquid = Volumetric Mass Transfer Coefficient x Driving Force

$$R_s = K_L \frac{A}{V}(C_s - C) = K_L a (C_s - C) \tag{3.2}$$

K_L = overall mass transfer coefficient (m/s)
A = total gas-liquid contact area(m^2),
V = volume of the liquid within the boundary(m^3),
a = specific contact area($1/m$),
= total contact area/volume of the liquid,
C_s = equilibrium aqueous concentration of solute(mg/L),
C = aqueous concentration of solute(mg/L)

Under unsteady-state, batch operation, the relation becomes

$$\frac{dC}{dt} = K_L a (C_s - C_t) \tag{3.3}$$

According to the two-film model of gas transfer (Skellard, 1974), the overall mass transfer coefficient may be influenced by both the liquid and gas-phase mass-transfer resistances

Total Resistance	=	Liquid-phase Resistance	+	Gas-phase Resistance

$$\frac{1}{K_L a} = \frac{1}{k_L a} + \frac{R_c T}{H_c'} \frac{1}{k_G a}$$

k_L, k_G = liquid- and gas-phase transfer coefficient(m/s),
R_c = universal gas constant (atm m^3/mol°K),
H_c' = Henry's constant (atm m^3/mol),
T = absolute temperature(K)

If we assume the gas phase and liquid phase have a common interface, Eq. 3.4 can be expressed as

$$\frac{1}{K_L} = \frac{1}{k_L} + \frac{1}{H_c k_G} \tag{3.4}$$

where H_c = dimensionless Henry's constant.

$K_L a$ can be determined by a batch procedure which is usually performed to determine the oxygen transfer capacity of aeration equipment. An unsteady state mass transfer model can be obtained by integrating Eq. 3.1

$$\ln\left(\frac{C_s - C_t}{C_s - C_0}\right) = -K_L a t \tag{3.5}$$

If the accumulation of emitted VOCs in the air surrounding the RBC unit is negligible, the equilibrium concentration of VOCs may be considered to be zero, thus Eq. 3.3 will be reduced to

$$\ln\left(\frac{C_t}{C_0}\right) = -K_L a t \tag{3.6}$$

For steady-state continuous flow operation, a simple mass balance of a VOC throughout the n equal-volume completely mixed regions can be written as:

$$\frac{C_n}{C_0} = \frac{1}{(1+(V_i/Q)K_L a)^n} \tag{3.7}$$

For plug-flow reactor, a steady-state equation can be written as:

$$\frac{C}{C_0} = e^{-K_L a \frac{V}{Q}} \tag{3.7a}$$

where Cn = solute concentration from the n-th stage, n = number of stages, and Vi = volume of a single stage. Thus, the $K_L a$ of a single completely mixed reactor or plug-flow

$$K_L a = \frac{C_0 - C}{C}\frac{Q}{V} \tag{3.8}$$

$$K_L a = \frac{Q}{V}\ln\left(\frac{C_0}{C}\right) \tag{3.8a}$$

Considering the two different mass transfer mechanisms, a convenient way to express the overall mass transfer rate coefficient, $K_L a$, can be written as:

$$K_L a = K_L a_l + K_L a_d \tag{3.9}$$

where K_La_l, K_La_d are the mass transfer rate coefficients for bulk water surface transfer and water film transfer on the disks. The fraction of the total K_La associated with either K_La_l or K_La_d depends on the operating conditions and the configuration of the system. Literature shows that mass transfer through the water film on the disk dominates under the usual operating conditions for typical RBC systems. The bulk water surface contributed less than 5% of the total contact area in our pilot system, and less than 1% of the total contact area for most of the commercial systems. Therefore, the mathematical model for physical VOC mass transfer considers the water film on disk only.

Eq.3.4 considers the overall resistance to interphase mass transfer as the sum of a gas-phase and a liquid-phase resistance. When the gas is only slightly soluble, Hc' is large, making the gas phase resistance negligible. Mackary (1975) recommended that the liquid-phase resistance dominates for the compounds having Hc' larger than 0.0048 atm m^3/mol. Oxygen and most of the VOCs satisfy this definition. The overall mass transfer coefficient, K_L, is then approximately equal to the liquid-phase mass transfer coefficient. The mathematical model presented in this chapter is mainly based on a liquid-phase resistance controlled mechanism and the following assumptions:

1. It is assumed that the water film on the disk surface is uniform and stagnant on the disk surface throughout the rotation; the velocity of the water film is equal to the rotational speed of the disks.

2. The VOCs' transfer from the water film to the gas phase is considered as a one-dimensional unsteady-state diffusion transfer process. Only the diffusion in the direction of vertical to the disk surface will be taken into account and those of other directions are neglected.

3. The distribution of VOCs in the water film is uniform at the point of emergence, and it is assumed that from then on the concentration of each solute on the surface of the water film is uniformly equal to its equilibrium concentration, Cs.

4. The VOCs concentrations are uniform throughout the basin owing to mixing and turbulence.

5. Since the concentrations and solubilities of VOCs and oxygen are very small, it is further assumed that the diffusion of solutes into/out of the water does not appreciably affect the temperature or other physical properties of the water.

Since the physical transfer of VOCs between air and

water occurs basically on the exposed disk surface, it is reasonable to treat the RDC as a compartmentalized system, including an aeration zone and an accumulation zone. In the aeration zone, VOCs in the water film attached on the disk surface transfer into the adjacent air and oxygen transfers from the air into the water film. In accumulation zone, VOCs transfer from the bulk water into the water film on the submerged disk and oxygen transfers from the water film into the bulk water. The molecular diffusion process between air and water films is predominant during the exposure of the disk to the air. In the latter step, while part of the disk submerges into the bulk water, VOCs are transferred from the bulk water into the water film by molecular diffusion as well as mixing.

Transfer of VOCs During the Disk Exposure

The exposure time of a small surface element, dA, on the disk is determined as:

$$t_a = \frac{2(\pi - \varphi)}{w} \tag{3.10}$$

$$\varphi = \arccos(R_i/r)$$

t_a = exposure time of element dA in air,
R_i = distance between bulk water surface and axis,
r = distance between element dA and disk axis.

If it is assumed that the edge effects of the disks can be neglected, the mass transfer of VOCs can be simulated by the diffusion mass transfer in a thin layer of water on a slab and equated as Fick's second law of molecular diffusion:

$$\frac{dC}{dt} = D\frac{\partial^2 C}{\partial x^2} \tag{3.11}$$

The initial and boundary conditions

$$\begin{array}{lll} t = 0, & 0 < x < \delta, & C = C_i, \\ t > 0, & x = \delta, & C = C_s \\ t > 0, & x = 0, & \dfrac{\partial C}{\partial x} = 0 \end{array}$$

where delta = thickness of the water film, Ci = initial solute concentration in the water film at the point of emergence, Cs = equilibrium concentration of the solute, D = diffusivity of the solute. Eq. 3.11 could be solved by either method of Laplace transformation or by the method of Separation of Variables (Skelland, 1974).

The solution by Laplace transformation (Crank, 1956) is

$$\frac{C_a - C_i}{C_s - C_i} = \sum_{n=0}^{\infty}(-1)^n \mathrm{erfc}\left\{\frac{(2n+1)\delta - x}{2\sqrt{Dt}}\right\} + \sum_{n=0}^{\infty}(-1)^n \mathrm{erfc}\left\{\frac{(2n+1)\delta + x}{2\sqrt{Dt}}\right\} \tag{3.12}$$

where,

$$\text{erfc}(x) = \frac{2}{\sqrt{\pi}} \int_x^\infty e^{-y^2} dy$$
$$= \text{complementary error function of } x \text{ (Carslaw, 1947)}$$
$$= 1 - \text{erf}(x)$$

The solution by Separation of variables is

$$\frac{C_a - C_s}{C_s - C_i} = \frac{4}{\pi} \sum_{n=1}^{\infty} \frac{(-1)^n}{2n+1} \cos\left\{\frac{(2n+1)\pi x}{2\delta}\right\} \exp\left\{\frac{-(2n+1)^2 \pi^2 Dt}{4\delta^2}\right\} \tag{3.13}$$

The series of Eq. 3.13 converges rapidly for large times. However the series of Eq. 3.12 converges more rapidly for small times. In a RDC system, the water film on the disk is exposed to the air from only a few seconds to thirty seconds. For such short diffusion times, Eq. 3.12 is of most value and adopted by most authors.

In a steady state continuous flow system, the concentration Ci is constant during the operation. Owing to the inherent nature of film transfer of the rotating disk, the diffusion mass flux between the air and exposed disk surface is time-depedent. The time-dependent mass flux of diffusing solute aross the air-water boundary layer at time t is expressed by Yamane and Yoshida (1972) as:

$$n_t = -D\left(\frac{\partial C}{\partial x}\right)_{x=\delta}$$
$$= -(C_s - C_i)\sqrt{\frac{D}{\pi t}} \left\{1 + 2\sum_{n=1}^{\infty} (-1)^n \exp\left(-\frac{n\delta^2}{Dt}\right)\right\} \tag{3.14}$$

The average mass flux of diffusing solute, N_t, over the exposure time, t_a, is

$$N_t = \frac{\int_0^{t_a} n_t dt'_a}{\int_0^{t_a} dt'_a}$$

$$N_t = 2(C_s - C_i)\sqrt{\frac{D}{\pi t_a}}\left[1 + 2\pi\sum_{n=1}^{\infty}(-1)^n \text{ierfc}\left(\frac{n\delta}{\sqrt{Dt_a}}\right)\right] \tag{3.15}$$

where

$$\text{ierfc}(x) = \int_x^{\infty} \text{erfc}(y)\,dy \quad \text{(Carslow, 1947)}$$
$$= \text{integral complementary error function of } x.$$

The liquid film mass transfer coefficient, k_L, can be equated as

$$k_L = \frac{|N_t|}{C_s - C_a^*}$$

where Ca^* is the effective average solute concentration in the liquid film. It is usually assumed that $Ca^* = Ci$. Thus, the average mass transfer coefficient for element dA is given by

$$k_L = \frac{\int_0^{t_a} n_t dt'_a/t_a}{C_s - C_i} \tag{3.16}$$
$$= 2\sqrt{\frac{D}{\pi t_a}}\left[1 + 2\sqrt{\pi}\sum_{n=1}^{\infty}(-1)^n \text{ierfc}\left(\frac{n\delta}{\sqrt{Dt_a}}\right)\right]$$

Mass Transfer in the Accumulation Zone

It is assumed that the accumulation zone is completely mixed with an uniform instant concentration of C_t. A mass

balance in the element dA at the point of re-submergence can be written as:

Residual Mass of Solute in dA	=	Initial Mass of Solute at the Point of Emergence	+	Total Mass Transfer During the Exposure

$$dA\delta C_{a,t} = dA\delta C_{i,t-ta} + dA\int_0^{t_a} n_t dt'_a \tag{3.17}$$

where $C_{a,t}$ and $C_{i,t-ta}$ are the solute concentrations in the water film at the points of submergence and emergence at time t and $t-t_a$ respectively. Most authors assumed that the water film was completely mixed with the bulk water. However, based on the oxygen transfer and tracer data obtained from flat disk, Zeevalkink (1978) and Matsuo (1985) suggested that the above assumption was not correct. Matsuo (1985) considered the only a fraction, f, of the water in element dA was interchanged with the bulk water during each revolution. Therefore, a mass balance in element dA at the emerging point can be written as:

$$dA\delta C_{i,t} = (1-f)dA\delta C_{a,t-t_w} + f dA\delta C_t \tag{3.18}$$

where t_w = retention time of element dA in the bulk water, and f ranged between 0 and 1 corresponding to no interchange and complete interchange. Combining Eq. 3.16, 3.17 and 3.18

$$\int_0^{t_a} n_t dt'_a = t_a k_L (C_s - C_{i,t-ta}) \tag{3.19}$$

$$C_{i,t} = (1-f)[(1-k)C_{i,t-T} + kC_s] + fC_t \tag{3.20}$$

$$k = k_L/(\delta/t_a), \quad \text{and} \quad T = t_a + t_w, \quad \text{time per revolution}$$

Since $K_L a$ and T are constants for each operating condition, $C_{i,t}$ can be equated as Eq. 3.21 for unsteady-state operations from Eq. 3.3, 3.4, by assuming

$$C_{i,t} = k' C_{i,t-T}, \qquad C_{i,t-T} = \frac{C_{i,t}}{k'}$$

$$C_{i,t} = \frac{k'[k(1-f)C_s + fC_t]}{(k'-1) + [(1-f)k + f]} \tag{3.21}$$

In steady-state continuous flow operation, the change of C_t is balanced by the bulk influent and effluent flows, thus

$$C_t = C_{t-T}$$
$$C_{i,t} = C_{i,t-T}$$
$$C_{a,t} = C_{a,t-T}$$

and k' is equal to unity. Since $K_L a$ and T are small for one revolution, the k' is still very close to unity in a batch operation. Eq. 3.21 is then simplified to Eq. 3.22 which has been derived from Matsuo (1985).

$$C_{i,t} = \frac{k(1-f)C_s + fC_t}{(1-f)k + f} \tag{3.22}$$

From Eq. 3.17, 3.18. the total interfacial mass flux of Eq. 3.16 can be equated as

$$\int_0^{t_a} n_t dt'_a = f(C_a - C) = t_a k_L (C_s - C_i) \tag{3.23}$$

The total mass balance of a diffusing solute in the bulk water for batch operation is expressed as Eq. 3.24 by combining Eq. 3.10 and 3.23.

$$\begin{aligned} V\frac{dC}{dt} &= \int_{R_i}^{R} f(C_a - C) w r dr \\ &= \int_{R_i}^{R} 2k_L (C_s - C_i)(\pi - \varphi) r dr \\ &= \overline{k}_L (C_s - \overline{C}_i) \int_{R_i}^{R} 2(\pi - \varphi) r dr \\ &= \overline{k}_L A_a (C_s - \overline{C}_i) \\ &= \frac{f}{(1-f)k + f} \overline{k}_L A_a (C_s - C) \\ &= (k_L)_{obs} A_a (C_s - C) \end{aligned} \tag{3.24}$$

in which $(k_L)_{obs}$ and k_L are related by

$$(k_L)_{obs} = \frac{f}{(1-f)k + f} \overline{k}_L \tag{3.25}$$

Three variables, exposure time, film thickness, and f are theoretically or empirically equated. The exposure time in this study, t_a, is simply calculated geometrically from the time per rotation, T, and the fraction of wetted band that is exposed to the air (Eq. 3.26, 3.26a). To evaluate the average exposure time, Yamane and Yoshida (1972) considered the mass transfer capacity and derived

Eq. 3.26b by using the relationship of Eq. 3.31. Bintanja (1975) derived Eq. 3.26c and Eq. 3.26d, corresponding to the case of Eq.3.31 and 3.32 respectively. A numerical comparison of Eqs. 3.26a, 3.26b, 3.26c, and 3.26d is shown in Table 3.1. Table 3.1 show that all four equations give very similar results.

$$t_a = T\frac{A_a}{A_a + A_w}$$
$$t_w = T\frac{A_w}{A_a + A_w} = T - t_a \tag{3.26}$$

$$t_a = \pi(R^2 - R_i^2) - \left[\pi R^2(\phi/\pi) - R_i\sqrt{(R^2 - R_i^2)}\right]\frac{1}{w}$$

$$t_a = \frac{1}{w}\frac{\left(1 - \left(\frac{R_i}{R}\right)^2\right) - \left(\frac{\phi}{\pi} - \frac{R_i}{R}\sqrt{1 - (R_i/R)^2}\right)}{1 - \left(\frac{R_i}{R}\right)^2} \tag{3.26a}$$

$$(w/60)\bar{t}_a = \frac{\left\{\int_{R_i/R}^1 \zeta\left[1 - \frac{1}{\pi}\arccos\left(\frac{R_i - R}{\zeta}\right)\right]^{1.5} d\zeta\right\}^2}{\left\{\int_{R_i/R}^1 \zeta\left[1 - \frac{1}{\pi}\arccos\left(\frac{R_i - R}{\zeta}\right)\right] d\zeta\right\}^2} \tag{3.26b}$$

$$(w/60)\bar{t}_a = \frac{\left\{\int_{R_i/R}^1 \zeta\left[1 - \frac{1}{\pi}\arccos\left(\frac{R_i - R}{\zeta}\right)\right] d\zeta\right\}^2}{\left\{\int_{R_i/R}^1 \zeta\left[1 - \frac{1}{\pi}\arccos\left(\frac{R_i - R}{\zeta}\right)\right]^{0.5} d\zeta\right\}^2} \quad \text{when} \quad \frac{\delta}{\sqrt{D\bar{t}_a}} \geq 1.7 \tag{3.26c}$$

$$(w/60)\bar{t}_a = \frac{\int_{R_i/R}^1 \zeta\left[1 - \frac{1}{\pi}\arccos\left(\frac{R_i - R}{\zeta}\right)\right] d\zeta}{\int_{R_i/R}^1 \zeta\, d\zeta} \quad \text{when} \quad \frac{\delta}{\sqrt{D\bar{t}_a}} < 0.8, \tag{3.26d}$$

$$\delta = \frac{4}{15}\sqrt{\frac{2\nu}{g}}\sqrt{wR\sin\phi} \tag{3.27}$$

$\phi = \arccos(R_i/R)$,
g = acceleration of gravity,
w = rotational speed (rpm),
ν = kinetimatic viscosity of water.

Since it is not possible to accurately measure the actual film thickness on the disk, the thickness is calculated from a theoretical equation. The thickness of water film on a flat disk, Eq. 3.27, was given theoretically by Zeevalkink (1978). Some other authors also proposed different equations to calculate the film thickness (Livich, 1968; Schliting, 1968). The disks used in this study were not flat disks. Because of the irregularities (roughness) and specific configurations of the disk surface a thicker liquid film will be actually retained on the disk surface than the prediction. The coefficient f in Eq. 3.28 was derived semi-theoretically by Matsuo (1985) on flat disks. In this study, the value of f still has to be empirically correlated with the physical parameters.

$$\begin{aligned} f &= f_1 + (1 - f_1)f_2 \\ f_1 &= \frac{\sqrt{Dt_w}}{\delta}\left(\frac{1}{\sqrt{Dt_w}} - \text{ierfc}\left(\frac{\delta}{\sqrt{Dt_w}}\right)\right) \\ f_2 &= \alpha\nu^{-0.3}w^{1.3}R^{2.1}(\sin\phi)^{1.5} \end{aligned} \tag{3.28}$$

$$K_L \sim k_L \tag{3.29}$$

$$K_L = \frac{k_L H_c k_G}{H_c k_G + k_L} \tag{3.30}$$

According to Eq.3.16, $\frac{k_L}{2\sqrt{D/\pi t_a}}$ can be represented graphically as a function of $\sqrt{\delta/Dt_a}$ (see Fig. 3.2).

For high rotational speeds, the length of diffusion is small compared with the thickness of the water film. The penetration model of Higbie holds (Bird, 1960). Therefore,

$$\frac{\delta}{\sqrt{Dt_a}} > 1.7 \quad \text{and} \quad \frac{\bar{k}_L}{2\sqrt{D/\pi t_a}} = 1, \qquad \bar{k}_L = 2\sqrt{\frac{D}{\pi t_a}} \tag{3.31}$$

For low rotational speeds, i.e. longer contact times, there is a high degree of water film saturation on the disk, Higbie's assumption of diffusion in a semi-infinite medium is no longer valid. Under ideal conditions, if the liquid film is fully saturated during the emerged cycle and completely mixed with the bulk water during the submerged cycle. Eq. 3.16 is reduced as

$$\delta / \sqrt{Dt_a} < 0.8 \quad \text{and} \qquad \bar{k}_L = \frac{\delta}{t_a} \tag{3.32}$$

Considering the situation that the water film is highly saturated (the penetration model does not hold) but not fully saturated or does not fully mix with the bulk water, a convenient way to express the liquid film mass transfer coefficient is

$$\bar{k}_L = f_E \frac{\delta}{t_a} \tag{3.33}$$

in which f_E is an effectiveness factor equal to the product

of the degree of mixing between bulk water and water film (Eq.3.25), fractional degree of saturation, and mass transfer enhancement factor. Rittmann (1983) also applied a similar expression for this region.

Data of oxygen transfer indicate that, for highly volatile compounds, Eq. 3.32 hold for normal operating conditions. Combining Eq.3.26a and 3.27 in accordance with Eq.3.32, k_L may be equated as Eq.3.34

$$\bar{k}_L = f_E \frac{4}{15}\sqrt{\frac{2\nu}{g}} w^{1.5} R^{0.5} F(\frac{R_i}{R}) \tag{3.34}$$

$$F\left(\frac{R_i}{R}\right) = \frac{\sqrt{\sin(\arccos(\frac{R_i}{R}))}\left(1-(\frac{R_i}{R})^2\right)}{1-(\frac{R_i}{R})^2 - \frac{1}{\pi}\left(\arccos(\frac{R_i}{R}) - (\frac{R_i}{R})\sqrt{1-(\frac{R_i}{R})^2}\right)}$$

It has been assumed in most studies that the mass transfer of VOCs or oxygen is controlled by the liquid film resistance. In this case, the overall mass transfer coefficient, K_L can be approximated by a liquid phase mass transfer coefficient (Eq. 3.29). It is true for highly volatile component, like oxygen. However, for those less volatile compounds, like chloroform, this assumption may not hold. For such cases, the overall mass transfer can be expressed as Eq.3.4 in accordance with the two film theory, and will be calculated by Eq.3.30.

Table 3.1 Numerical Comparison of Eqs. 3.26a, 3.26b, 3.26c, 3.26d

	w/60 t_a			
Ri/R	Eq. 3.26a	Eq. 3.26b	Eq. 3.26c	Eq. 3.26d
0	0.5	0.5	0.5	0.5
0.1	0.559	0.561	0.560	0.559
0.2	0.611	0.615	0.612	0.611
0.3	0.657	0.662	0.658	0.657
0.4	0.700	0.704	0.701	0.699
0.5	0.739	0.743	0.740	0.739
0.6	0.778	0.780	0.778	0.777
0.7	0.816	0.818	0.816	0.817
0.8	0.855	0.856	0.855	0.855
0.9	0.902	0.902	0.901	0.901

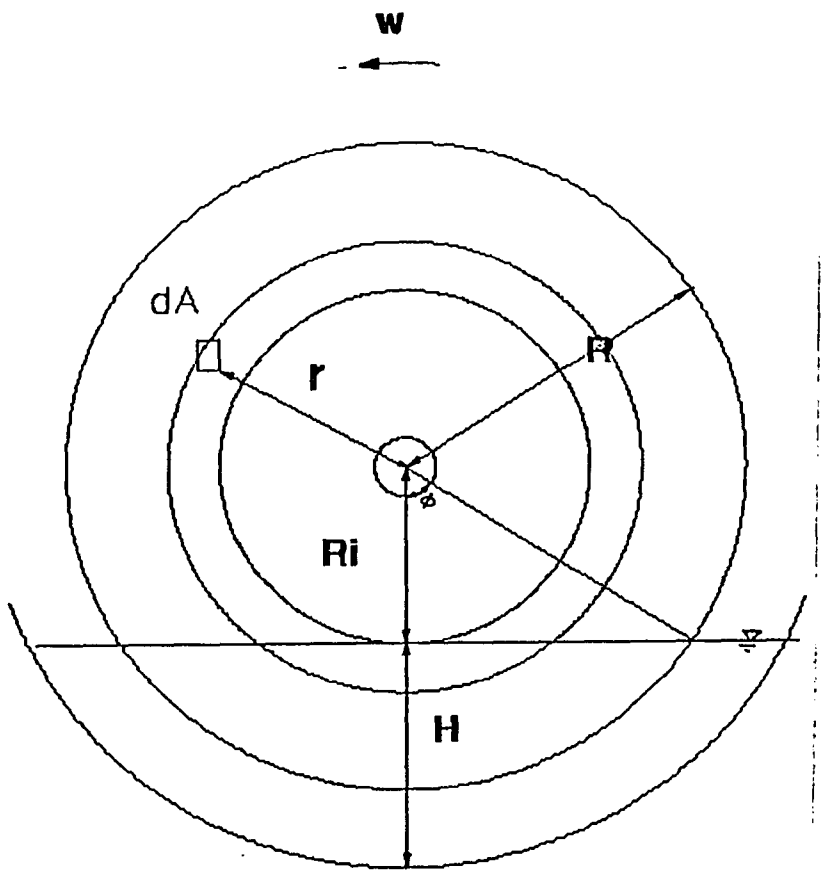
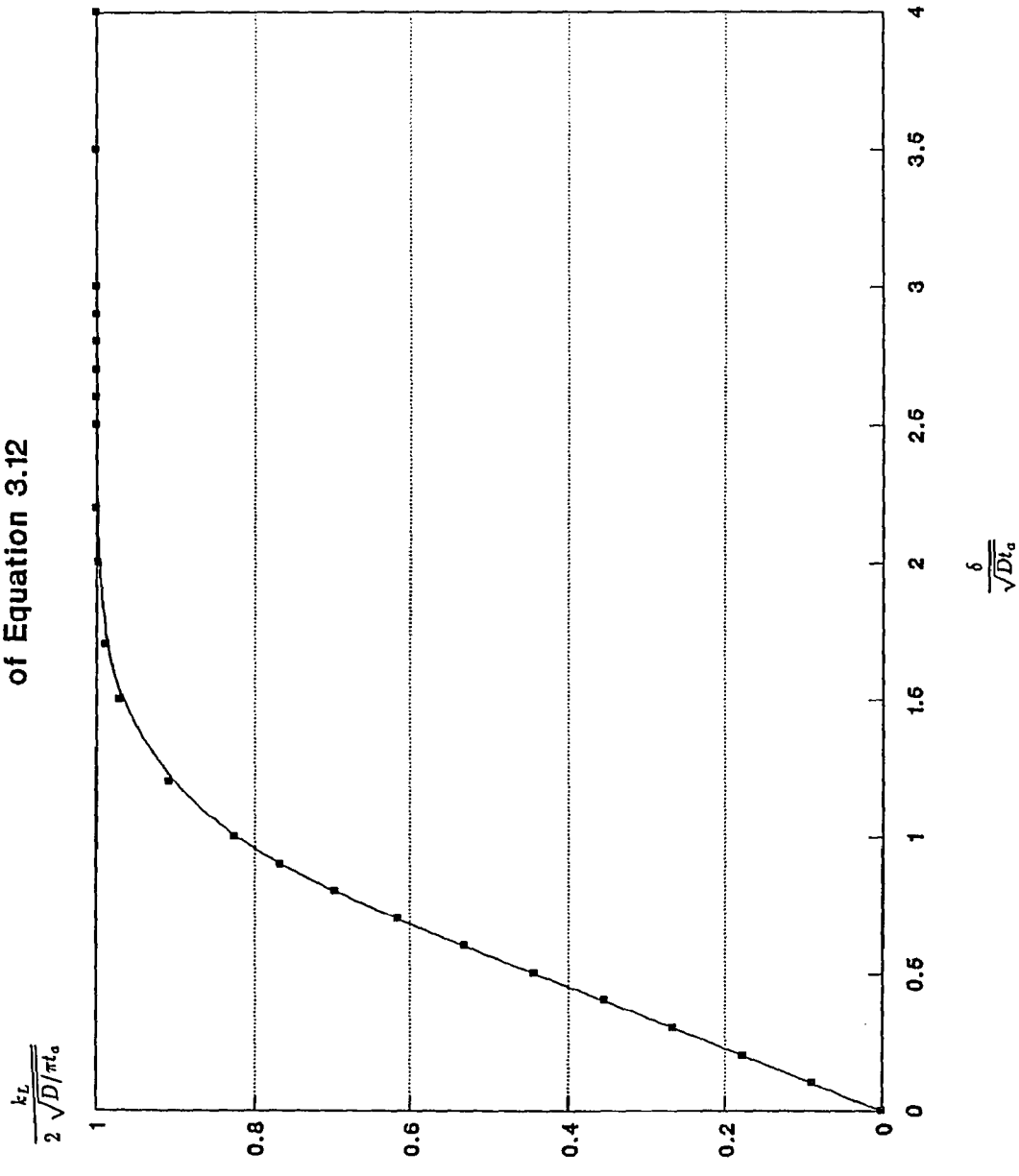


Figure 3.1 Schematic Diagram of the RDC Unit

Fig 3.2 Graphical Representation of Equation 3.12



CHAPTER 4 MATERIALS AND METHOD

4.1 Rotating Biological System

A laboratory-scale, continuous-flow, rotating disk contactor pilot plant was used for the present study to treat a synthetic wastewater spiked with selected volatile organic compounds. A schematic flow diagram of the RBC system is shown in Figure 4.1. The RDC was manufactured by Lyco Co., and consisted of four basins in series, with six discs (38 cm in diameter) for each basin, which allowed the system to be operated as a four-stage reactor or as a single-stage reactor with four compartments. The disks were not flat disks, but with sockets of 1 inch in diameter evenly distributed on the surface. The total surface area of media in each basin was 20,000 cm^2. Compared with flat disks of the same diameter, the disks used in this study provide about 1.46 times the area of flat disks. The maximum wetted volume of each basin is equal to 6.0 liters. The dimensions of the pilot plant are listed in Table 4.1.

All parts in contact with the wastewater were made of Teflon or metal to minimize any possible contamination. For the single-stage, continuous-flow study, the synthetic wastewater was fed to the basins of the RBC system from four inlets by a Masterflex .pa pump, equipped with speed

controller, at the predetermined rate.

Complete mixing in the basin was obtained by vigorously stirring of the rotating discs. The tracer technique using sodium chloride was applied to determine flow patterns in the basins under different rotational speeds. Initially the basin of the RDC was filled with demineralized water. A sodium chloride solution of 1,000 mg/l was then pumped into the reactor, resulting in average hydraulic retention times ranging between 10 to 60 minutes. The basin was agitated by disc rotation at rotating speeds of about 1, 1.6, and 13 rpm respectively. The tracer, sodium chloride, was monitored with a conductivity bridge (YSI conductivity bridge, Model 31). The operating conditions were similar to those used in this study. The good agreement between the experimentally determined response curves and the theoretical curve ensures that the RDC was operated under a completely mixed condition in most of the experiments. A detailed discussion will be presented in Chapter Five.

4.2 Synthetic Wastewater

The synthetic wastewater was prepared by dissolving trichloroethylene (TCE), tetrachloroethylene (PCE), ethylbenzene (EB), chloroform (CLF), and 1,1,1-trichloroethane

(TCEN) in tap water. These five compounds are chosen because their Henry's constants covered a range of nearly 3 orders of magnitude and they are frequently observed in wastewater and groundwater. The feed solution was prepared freshly for each run. The final concentrations of VOCs were ranged from 200 ug/l to 5,000 ug/l corresponding to the concentrations in most contaminated ground water or wastewater. Physical-chemical properties of these six compounds are listed in Table 4.2.

All the chemicals were obtained from Aldrich Chemical Co. (Milwauee, WI). A 1,000 mg/l stock solution was prepared by dissolving 1 g of each test compound in 1 l of methanol. Methanol is water-miscible and biodegradable, serving as a cosolvent to facilitate the preparationof solution. Previous studies indicated no mutual effects on the Henry's constants of trichloroethylene, chloroform, tetrachloroethylene, 1,1,1-trichloroethane, and dichloro-methane in a aqueous solution of the five at up to a total mixture concentration of 375 mg/l (Gossett, 1985). Gossett (1987) reported no impact of methanol on the apparent Henry's constant of TCE at up to 2% (v/v) methanol, while a 5% methanol solution depressed Henry's constant by about 7%. He also studied the effects of ionic strength of up to 1.0 M KCl on Henry's constants of the above five compounds. Among these five compounds,

tetrachloroethylene exhibited the greatest ionic strength dependence, and ionic strength would have to exceed 0.2 M (KCl) to cause a greater than 10% increase in Henry's constant. He concluded that the practical impact of ionic strength had appeared to be minimal. In this study, only less than 0.5% stock solution was spiked into tap water to make the synthetic wastewater. The influences of methanol and dissolved salt are considered insignificant.

4.3 Operation

The experiments were conducted in a well ventilated room. The RDC pilot plant was placed nearby a forced-draft exhausting fume hood to prevent accumulation of organic compounds in the air surrounding the pilot plant. The overall experimental procedure was divided into two operations: batch and continuous operations. The effects of rotational speed (1.5 to 18 rpm), depth of immersion, distance between adjacent disks, VOC concentration and temperature (15 to 30 ^{o}C) on VOCs stripping efficiency were tested by the batch reactor, non-steady state procedure. Since previous mass transfer studies concentrated on the oxygen absorption, it was used as a reference compound in this study for comparison. Therefore, in addition to VOCs, oxygen concentration was also measured. The air stripping efficiency, and effects of hydraulic retention time,

staging, and air flow rate were evaluated by continuous operation under constant bulk water volume but varied disk rotational speeds.

4.3.1 Batch Operation

The synthetic wastewater for the experiments was made of tap water and methanol-VOCs solution. The oxygen was first purged from the water by bubbling nitrogen through the water. Then, 5 l of deoxygenated water was spiked with 1 to 5 ml stock solution, resulting in concentrations on the order of 0.2 to 5 mg/l. The feeding bottle was capped, sealed and mixed well with a magnetic stirrer. The synthetic wastewater was then stored under constant temperature for at least 12 hours to allow the solution been equilibrated to a constant temperature. After the storage, the desired quantity of synthetic wastewater was added into the RBC unit. Following 2 min. mixing through the rotation of the discs, an initial sample was taken. Thereafter samples were taken at regular time intervals. Concurrently, the oxygen concentration in bulk water was measured with a dissolved oxygen probe. The VOCs were analyzed using gas chromatography and purge-and-trap methods.

The effects of rotational speed, depth of immersion,

and water temperature were evaluated by varying the level of one parameter while the others remained constant. The immersion depth of disk was changed by altering the volumes of the bulk water and hence the water surface level. The physical characteristics of the system under different bulk water volume shall be seen in Chapter Five. The water volume ranged between the maximum reactor capacity of 6 L to 1 L. At least five runs of different rotational speeds were conducted for each immersion depth. Since rotational speed is one of the most important operating parameters, more runs were conducted with 4 L bulk water to observe the effects of different rotational speeds in the range of 1 to 18 rpm. The effect of water temperature, 10 to 35^{o} C, was studied under various rotational speeds with 4 L of bulk water in the trough.

The wetted area of the discs was determined by directly measuring the radial width of the wetted band with a scale. In a series of experiments at different rotational speeds, it was found that the wetted discs' area was independent of rotational speed and is only a function of immersed depth. This result confirmed many other observations in the literature (Bintanja, 1975). The area of the free liquid surface at the bulk solution was less than several percent of the total wetted area. Thus, it was not included in the total contact area. The total

contact area was therefore defined as the total wetted area of discs that was above the free water surface. Most of the authors assume that the wetted area of the disks is equivalent to the contact area. The specific area was then calculated by dividing the values of wetted area by the volume of bulk water.

4.3.2 Continuous Operation

The determination of stripping efficiency in the continuous flow study was conducted under a condition similar to that of the batch study except a continuous feed was applied to the system. Two types of reactor systems were tested: four basins in series (four stages), four basins in parallel (single stage). The experiments were conducted under ambient room temperature with flow rate and rotational speed as the two operational parameters. The influent synthetic wastewater was equilibrated to room temperature before being pumped into the system. However a linear decrease of 2 to 4^{o} C was observed consistently across the system. This temperature decrease seemed to be independent of rotational speed and hydraulic loading and was attributed to the heat loss due to evaporation. The wet volume of the reactor was kept at about 6 L throughout the continuous flow study. Since the loss of VOCs from the storage tank by volatilization was inevitable, the decrease in VOC concentrations in the storage tank were monitored

every hour until the representative samples of the experiment were taken.

4.4 Analysis of Selected VOCs

4.4.1 Apparatus and Materials

1. Vial with cap: 40-ml, 10-ml, and 2-ml capacities, screw cap. Detergent washed, rinsed with tap and deionized water, and dried at 105^{o}C for 30 minutes before use.

2. Septum: Teflon-faced silicone. Detergent washed, rinsed with tap and deionized water, and dried at 105^{o}C oven for 30 minutes before use.

3. Gas chromatograph: Hewlett Packard 5890, with FID.

4. Concentrator: Tekmar 4000 Dynamic Headspace Concentrator.

4.4.2 Reagents

1. Organic-free water: deionized water, generated from water purification system.

2. Stock standard alcoholic solutions: prepared from pure standard chemicals and methanol. Stored at 4^{o}C, protected from lights, and replaced after 2 weeks.

3. Secondary dilution standards: prepared in methanol by mixing stock standard solutions together, stored at 4°C with zero headspace.

4. Calibration standards: prepared daily from secondary standards, discarded after 1 hr.

5. Quality control check standards: purchased from Supelco Co., which were used to determine the accuracy of calibration standards.

4.4.3 Analysis Method

U.S.EPA approved and recommended procedures for volatile organic compound analysis include purge-and-trap, closed-loop stripping and liquid-liquid extraction techniques. In this study, VOC concentration in liquid samples were analyzed according to EPA Method 601 purge-and-trap procedure. The purge-and-trap technique requires only a small sample (5 ml). It provides fairly high sensitivity and rapid, accurate and reproducible results. However, the transfer of the liquid sample from the original sampling tube to the stripping vessel will cause the potential loss of volatile constituents. It is important to transfer the sample as quickly as possible to minimize the exposure of the sample to the atmosphere. Duplicate samples were collected in two bottles separately. In order to minimize volatilization during sample collection, the sample bottles

were filled with care to avoid turbulence and bubbling. The sample bottle was filled to overflowing before being capped to remove the air headspace. Usually, collected samples were analysized immediately after collection. When this was not possible, samples were stored in a 4^{o}C refrigerator and analyzed within one or two days. The chromatographic conditions and settings for purge and trap procedure are presented in Table 4.3.

The GC was calibrated immediately prior to conducting the analysis and rechecked every four hours in each analysis session. The external calibration standards, mixtures of the test VOCs were prepared fresh daily at a minimum of four concentration levels, with the whole range bracketing the expected concentrations of each VOC. The calibration curve and calibration factor (CF) were obtained through analyzing the standards. The concentrations of the target compounds were calculated as:

$$CO = CF \times A \times PA/PB$$

where

CO = the concentration of the target compound

CF = calibration curve

A = area of the peak of target compound

PA = a constant, been set to 1

PB = a constant, been set to 1

For very low VOC concentration samples (20 ul/l and less), an internal standard calibration procedure was used. Components of known concentration were spiked into the external standards or wastewater samples. The spiked samples were then analyzed by GC. The concentration of target compound was calculated as:

$$CO(T) = CT \times A(T)/A(I) \times PB/PA$$

where

CO(T) = concentration of target compound
CT = slop of A(I)/A(T) vs CO(T)/CO(I) calibration line
CO(I) = concentration of internal standard
A(T) = peak area of target compound
A(I) = peak area of internal standard

4.4.4 Quality Control

The QA/QC procedure employed in this study was basically followed the QA/QC program in EPA Method 601. Quality control check standards, purchased from Supleco Co. instead of from EPA's lab were used to determine the accuracy of the analysis. A distilled water blank was analyzed at the begining of an analysis session and at each time a set of samples had been extracted or a change in

reagent as a safeguard against chronic laboratory contamination. Replicates were analyzed to validate the precision of the analysis. Fortified samples, which were carried through all stages of sample preparation and measurement, were analyzed to validate the sensitivity and accuracy of the analysis. Spiked reagent water and wastewaters were analyzed to calculate the recovery (P) for each component as 100(A-B)%/T, where A was the analytically determined concentration, B was the mean background concentration, and T was the known true value of the spike compound. The recovery of each test compound is shown in Table 4.4.

Table 4.1 Dimensions of Rotating Disc Pilot Plant

Number of Stages	4
In each stage:	
length of trough	15 cm
width of trough	40 cm
number of discs	6
diameter of discs	38 cm
thickness of discs	0.15 cm
total area of discs	20,000 cm^2
distance between outer disc and trough	3 cm

Table 4.2 Properties of Compounds Studied

Compounds	Formula	Molecular Weight (gm/mol)	Henry's Constant (atm)	Henry's Constant (dmlss) [a]	Diffusivity [b] (10^{-9} m^2/s)	Solubility [e] (mg/l)
Trichloroethylene	$CHCl=CCl_2$	131.4	550	0.41	0.86	1100
Tetrachloroethylene	$CCl_2=CCl_2$	165.8	1100	0.82	0.79	140
1,1,1-Trichloroethane	CH_3CCl_3	133.4	400	0.30	0.83	720
Chloroform	$CHCl_3$	119.4	170	0.13	0.94	8200
Ethylbenzene	$C_2H_5C_6H_5$	106.2	358	0.27 [d]	0.90	152
Oxygen	O_2	32.0	43000	32.10	2.03	9.09

a. H(atm) = H (dmlss) * 4.57 * T (K)
   H(atm) = H (m atm/mole) * 55,60

b. Robert (1983)

c. Kavanaugh (1985)

d. Petrasek (1983)

e. EPA (1983)

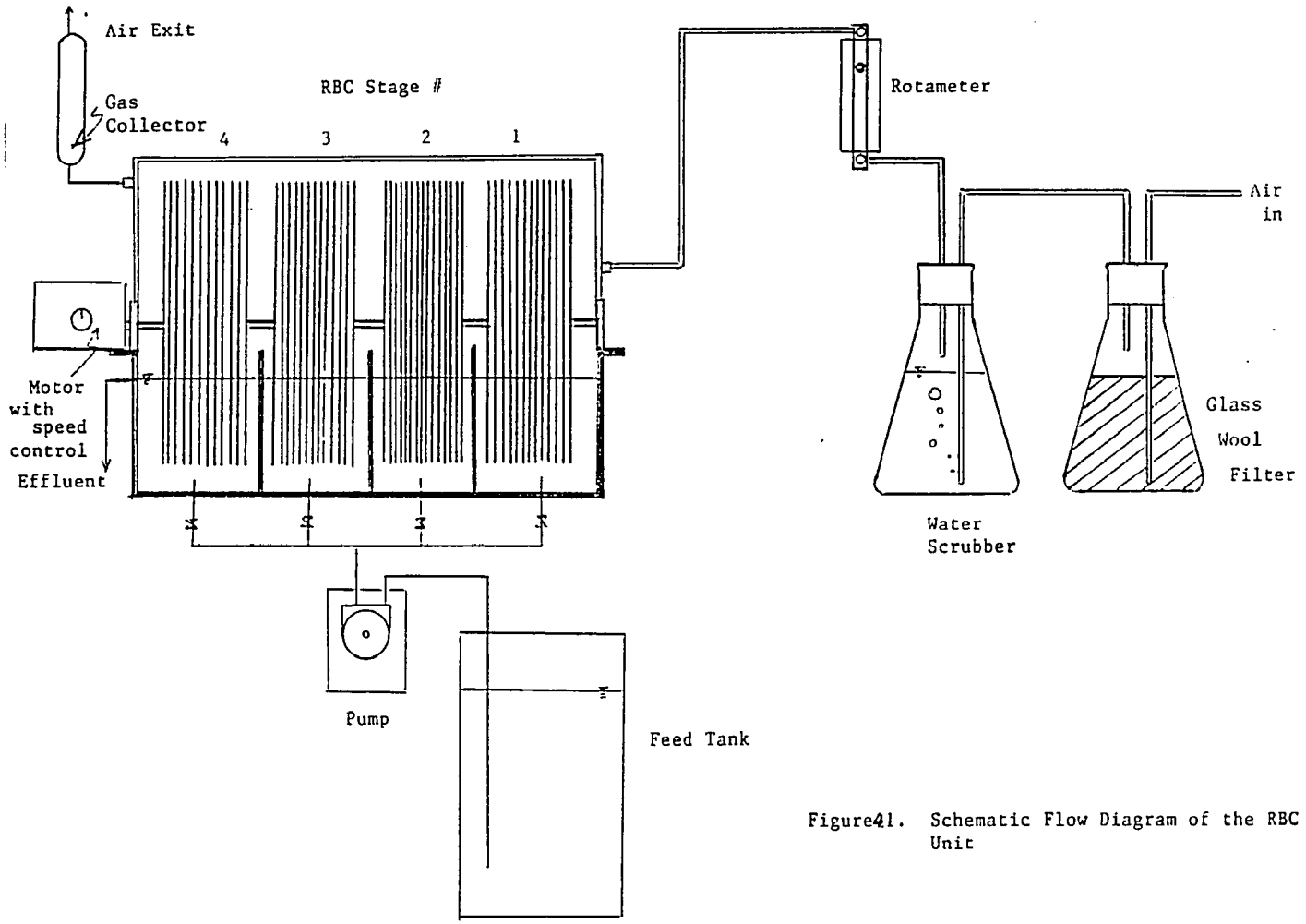


Figure41. Schematic Flow Diagram of the RBC Unit

Table 4.3 Conditions Used for the Analysis of VOCs

Gas Chromatography-Hewlett Packard 5890 GC	
Column	Supelco 60/80 Carbopack B/1% SP-1000
Detector	FID, 250° C.
Carrier Gas	Nitrogen, 40 cc/min.
FID Gas	Air 300 cc/min, Hydrogen 30 cc/min.
Injector	200° C.
Oven Program	45° C-12° C/min-200° C.
Purge-and-Trap	
Purge	11 min. at 30°C with Nitrogen (40 cc/min.)
Desorb	4 min. at 180°C.
Bake	10 min. at 225°C.
Sample Size	5 ml.

Table 4.4 Recovery of Analysis

Compounds	Recovery %
Chloroform	95
1,1,1 Trichloroethane	94
Ethylenebenzene	95
Trichloroethylene	92
Tetrachloroethylene	92

CHAPTER 5 RESULTS AND DISCUSSION

5.1 Parameter Estimation in Batch Study

Experiments to determine K_La were carried out without biomass in the RDC pilot system. Five volatile compounds: trichloroethylene, tetrachlor-ethylene, ethylbenzene, chloroform, 1,1,1 trichloroethane were used, in addition to oxygen, as the targets. Eq. 3.3 was the fundamental equation used to describe the absorption and desorption processes. This equation has been proved to fit adequately the vast majority of liquid-gas mass transfer data obtained from surface, submerged aeration tests and other interphase mass transfer studies. Values of the overall mass transfer rate constant (K_La) in this study were determined by either batch procedure or continuous flow procedure as described in Chapter 3. The batch procedure was similar to the unsteady state clean water oxygen transfer test which is widely applied to determine the oxygen transfer capacity of different aeration equipments. A variety of graphical and numerical procedures have been proposed to interpret the experimental data. Three of them, direct (differential) method, log deficit method, and exponential method, have been widely used to evaluate mass transfer efficiency. These three methods fit Eq. 3.3, 3.5 and 3.5a to the experimental data respectively (Baillod, 1979; Brown 1979).

The log deficit method was used in this study to calculate K_La values from the data obtained from batch studies. This method has long been the industry standard for determining the oxygen transfer rate (WPCA, 1971). In this study the experimental data sets for oxygen transfer were correlated as ln[(Cs-Ct)/(Cs-Co)] vs. time in accordance with equation 3.5. According to Eq. 3.5, a semilog plot of the deficit vs time should show a straight line with the slope corresponding to its K_La value. To determine the mass transfer rate of VOC from experimental data, Eq. 3.5 was modified to Eq. 3.6 by assuming Cs = 0. The data for VOC transfer were then correlated as ln(Ct/Co) vs time in accordance with the equation. Linear regression curve fitting techniques were used to draw the line through the plotted data. Several factors associated with the adequacy and accuracy of the test and the data analysis method are discussed in the following section.

The log deficit method has the advantage of using linear least squares for parameter estimation, but requires a given Cs value. The error in the determination of Cs will cause data points to distribute in other than a straight line on the semilog plot (Gilbert, 1976, Campbell, 1976). If the used Cs value is higher than the actual value, the plotted line tends to curve upward as the solute concentration approaches its equilibrium concentration.

When the used value is lower than the correct value, the line tends to curve downward. The Cs values can be obtained either by measurement, by a book surface saturation value, or by estimation from best fit analysis (Campbell, 1976). The Cs of oxygen used in this study was obtained by measuring the dissolved oxygen concentration in the trough after a 2-hour aeration period. If this measured value did not give a satisfactory straight line on the semilog plot of deficit, a value estimated by the Best Fit Log Deficit Method was then used instead (Campbell,1976). By this method, the value of Cs is estimated through a succession of regression analyses to obtain either the best straight line or the minimum residual sum of squares.

Although possible errors induced in measuring or estimating Cs may bias the estimate of $K_L a$, it appears in literature that the log deficit method usually yields results close to those obtained using much more sophisticated analysis methods (Ewing, 1979, Gilbert, 1976, 1977). Significant errors induced in Cs determination in the oxygen transfer test were mainly found in submerged aeration systems but not in surface aeration. Because partial pressures of the solutes in the air bubbles change with depth in a submerged aeration, Cs varies with depth and time during the unsteady state test. It is more difficultto determine the representative value of Cs forthe submerged aeration system. However, RDC is basically a

surface aeration system. Cs is constant and equal to the equilibrium concentration attained at "infinite" time, 2 hours in this study.

Since the batch procedure for open space air stripping is an unsteady state operation, VOC concentrations in the air and water continuously change during the test period. It is not possible to maintain a specific Cs value throughout the whole run. However, owing to the rapid replenishment of air by the exhausting fume hood nearby the RDC unit, the accumulation of VOCs in the air is implicitly ignored, and thus, the equilibrium VOC concentration in the water film is virtually negligible. Some typical data sets for this study are shown in Figure 5.1 to 5.5 as examples. The linearity of the data set attests to the goodness of fit of Eq. 3.6, and the appropriateness of applying the log deficit method for data analysis. Data from other studies of air stripping by surface aeration also gave the same conclusions (Robert, 1983). However, if the continuous instantaneous replacement of the air in the internal media voids is not achieved, the assumption of Cs = 0 may not hold. In this case, the K_La value calculated by Eq.3.6 will be less than the true K_La value, since Eq.3.6 considers the maximum driving force. In this study, Eq. 3.6 was further confirmed by the exponential method. However, if the results show lack of agreement with the

assumption, Eq. 3.5 or 3.5a may have to be applied instead.

In order to further test the assumption and appropriateness of the log deficit method, direct method and exponential method were also applied to analyze some of the experimental data as comparisons. The direct method has the advantage of using least squares analysis to evaluate experimental data without the need to assume a value for Cs. This method plots dCt/dt against concentration yielding a straight line on rectangular coordinates with the slope equal to the negative $K_L a$ value. The data used in Fig 5.1 were plotted in Fig 5.6 by direct method. The y-axis intercept is the maximum rate of oxygen transfer, $K_L aCs$. The x-axis intercept is the saturation concentration. The main disadvantage is that the direct method magnifies the noise in the data (Gilbert, 1976, 1977, Brown, 1979, 1982). Figure 5.1 shows that the log deficit method tends to smooth irregularities in the data resulting in nearly straight lines on the semilog plots. However, Figure 5.6, plotted by the direct method with the same sets of data shows noticeable fluctuations, even if the original experimental data are smooth (Fig 5.7). It is due to the numerical approximation of dCt/dt by taking differences between successive concentration values, that results in an error substantially larger than the error induced from concentration measurements. Stukenberg (1976) claimed that those flunctuations revealed surging in the mixing flow

pattern of the aerator. Therefore, the direct method can also be used to predict the flow pattern around an aerator in addition to $K_L a$ determination. since the estimates of $K_L a$ by direct method are more imprecise than the other methods, it was not used in this study for parameter estimation.

The exponential method fits Eq.3.5a to the experimental data for both air stripping of VOCs and absorption of oxygen. This equation mathematically describes the concentration response curve, C versus t (see Fig 5.7). The least-squares nonlinear regression analyses provided by Marquardt (1963), and Reed (1931) can be performed to estimate the unknown parameters, Cs and $K_L a$, assuming the value for any parameter. This estimation procedure includes iterative trial and error searches for the least squares estimates. The exponential method gives results usuallycloser to those of the Best Fit Log Deficit Method. A BASIC program provided by Baillod (1983) was used here to perform nonlinear regression analysis of the unsteady state mass transfer data.

It is evident that all the analysis methods have difficulties with the error structure of the dependent variables (Brown, 1982). Since the exponential method is more sensitive in the region of high transfer rate, it may

overweigh the observations at the beginning of the experiment where transfer rates are high. Due to the nature of logarithmic transformation and ability of experimental measurement, the log deficit method is more sensitive in the region of low deficit. Thus, in log deficit method, the error in the deficit increases as deficit decreases. This situation was noticed when dissolved oxygen was approaching its saturation value or when VOC concentration was reaching very low level.

In order to avoid taking logarithms of zero or negative deficit, the log deficit method usually requires truncation as solutes approaching saturation. Several authors recommended that concentrations observed at times greater than approximately 95 per cent should be truncated (Brown, 1982). Truncation increases the correlation between K_La and Cs but reduces the precision of their estimated values (Brown, 1979). It was observed from this study that truncating data in the vicinity of equilibrium was sometimes inevitable for oxygen transfer tests but not usually necessary for VOCs stripping tests. Such a result was attributed to the lower detection limit and higher precision of GC than of the oxygen probe.

Initial mixing of the solution and other physical test conditions may cause unreliable data generation at the beginning of the test. The data lower than 10 to 20 percent

saturation have been traditionally truncated in aeration tests. As it has been shown by Boyle (1974) that data at the beginning of the test have a small effect on the ability to estimate $K_L a$ and Cs. It is especially true when applying the log deficit method for data analysis. Owing to the nature of semilog plot, the data at the beginning of the test usually show less influence on the best fit line while the data at the end of the test have greater ability to influence the slope of the best fit line (see Fig 5.1 to 5.5). Conversely, the plot of concentration versus time in exponential method may overweigh the concentration changes at the beginning of the test. Therefore, it is a general practice of unsteady-state tests to use only the data between 20 and 80% saturation for analysis.

The unsteady state oxygen transfer tests should be conducted for as long a period of time as practicable. The minimum period of time should be longer than approximately $4/K_L a$ which corresponds to a dissolved oxygen concentration of 98% of Cs. It is also recommended that the regression analysis be based on a minimum of 10 to 15 data values. Approximately 2/3 of the data should be evenly distributed over the period of time between zero to $2/K_L a$ (Brown. 1982). These criteria were followed in this study for both the oxygen absorption and VOC stripping tests. The first sample was taken 1 minute after the initial mixing and was

used as the C_o value. After that, samples were taken at 30-seconds time intervals during the first 5 minutes, at 1-minute intervals during the second five minutes and then at 2-minutes intervals until the oxygen reached saturation or VOCs reached minimum detectable concentrations. Total sampling period was about 15 min, and 10 to 15 samples were obtained for each run. The dissolved oxygen concentrations above 95% saturation were truncated but none of the data at the beginning of the test was truncated.

The values of K_La and Cs obtained under various batch operating conditions are summarized in Table 5.1 and 5.2. A comparison of the results of the log deficit method and of the exponential method for some data sets is listed in Table 5.3. It shows that both methods gave very similar results for VOCs stripping and oxygen absorption. As mentioned above, variations of the results by the log deficit method were primarily associated with the selections of representative mean Cs. Since the RDC pilot plant used in this study was basically a surface aeration system with relatively small dimensions, the Cs would be constant and uniform throughout the bulk water. Similar results came out from both methods as expected. Since the log deficit method had been used by most of the unsteady state mass transfer studies found in literature, this method was applied for data analysis in this study to make the results comparable with those from other studies.

5.2 Parameter Estimation in Continuous Flow Study

Eq. 3.7 and 3.8 were developed on the assumption of a completely mixed flow pattern occurring in the RDC unit. To examine the correctness of the assumption, a tracer technique was applied to determine the flow pattern in each stage. As mentioned in Chapter 4, a sodium chloride solution with constant concentration of 1,000 mg/1 was continuously pumped to the RDC system until the concentration in the effluent was indistinguishable from that in the feed. The experiments of the tracer study were conducted with constant water volume of 6L and with a feed parallel to the shaft. Three hydraulic loading rates and three disk rotational speeds were tested. The normalized step input response curves, F-curve, of an ideally mixed tank, and the experimentally determined response curves are presented in Figure 5.8. Good agreement between the actual data plots and theoretical curve shows that nearly complete mixing characteristics were exhibited in individual stages for a wide range of flow rates and rotational speeds. Such a result demonstrated that each stage of the pilot RDC unit behaves like a CFSTR. Even at low Reynolds number, the mass transport in bulk water is mainly induced by turbulent convection, and uniform substrate concentrations exist in the bulk water. The whole 4-stage system may be treated as

a system of four CFSTR in series. The tracer study provided a good basis for developing Eq. 3.7 and 3.8. Similar results have been reported in both baffled pilot plants and full scale systems with flow perpendicular to the shaft (Watanabe, 1978; Olem, 1980; Larry, 1984, Shieh, 1982). Olem (1980) applied the procedure proposed by Pintenich (1979) to quantify the hydraulic characteristics of a continuous flow RDC unit, and reported that the flow pattern of a pilot plant with disks of 0.5 m in diameter showed only a slight deviation from a baffled full scale RDC (2 m in disk diameter) unit. However, there is no available data to demonstrate that an unbaffled full scale system with parallel feed will show similar flow characteristics.

The continuous flow air stripping was conducted under room temperature with a constant water volume of 6 L in each stage. A linear decrease of temperature of 2 to 3^{o}C was found consistently across the four-stage system. Such a temperature drop seemed to be independent of rotational speed and influent flow rate, and was attributed to evaporative cooling. The overall removal and individual removal efficiency of each stage obtained under various feeding rates and disk rotational speeds are presented in Table 5.4. The equivalent performances of different stages, except the first stage, implied the physical and

hydraulic similarities between each other. The first stage usually showed slightly higher removals than the others, and might be attributed to the extra loss at the entrance region and water transport in the tubing.

Some typical data sets are plotted in Fig 5.9, 5.9a, and 5.9b as examples. The scattered points in Figure 5.9 represent actual removal efficiencies. The solid lines represent the theoretical removal curves of a single and a four-stage completely-mixed reactor, which were calculated with the average removal efficiencies of the four stages. The four-stage RDC pilot unit behaved like a series of completely mixed treatment basins. With similar hydraulic retention times in each stage, the efficiencies increased with increasing number of stages. In each single stage, the removal efficiency increased, as expected, with increasing hydraulic retention time and rotational speed (see Figure 5.9a and 5.9b). The mass transfer rate constant, $K_L a$, was calculated in accordance with Eq. 3.7 and 3.8 for each volatile compound. To examine the influence of hydraulic retention time on the rate constant, the $K_L a$ values of TCE obtained from different hydraulic retention times were plotted in Fig 5.10 versus the disk rotational speed. There is no significant difference in $K_L a$ values, as shown in Figure 5.10. Considering all these observations, it was concluded that $K_L a$ is relatively independent of hydraulic retention time and is a function of disk rotational speed.

The behavior of an RDC system can be simulated by the system of a number of completely mixed reactors in series. It is appropriate to apply Eq. 5.1 to 5.8 to interpret the experimental data obtained from batch or continuous flow experiments.

As shown in Fig 5.9, with similar operating conditions, the four-stage RDC system consistently showed higher removal efficiencies than the single stage system. Such results agree with the general principles of reactions in mixed reactors. It is well known that for concentration dependent reactions, the required total volume of a series of mixed reactors is less than that of a single mixed reactor to achieve the same level of removal. The required volume of plug flow reactor is theoretically the smallest. Staging is usual practice in RBC systems. An RDC unit of four stages per shaft is not unusual for some commercial units. The arrangement of stages and flow is to uniform the organic and hydraulic loadings. However, in order to increase the driving force for air stripping, the flow in RDC should be arranged as close to a plug flow reactor as possible, thus, a parallel feed and more stages per shaft is preferred for the purpose of air stripping of VOCs.

5.3 Calculation of Specific Interfacial Area

Specific interfacial area is one of the important factors to determine the overall mass transfer rate. The effective interfacial area for oxygen absorption and VOCs desorption in the RDC system includes the liquid film on the exposed disk surface, rivulets, drops dripping from the disks, and the free surface of the bulk water in the trough. It is usually believed that the rivulets and the drops have only a minor contribution to the mass transfer rate in a RDC system. Thus, the RDC system basically provides two major interfacial surface areas. The relative importance of each interfacial area depends on the operation and the configuration of the system. The rotating disk aerators supplied for oxidation ditches and aerated lagoons (e.g. PFT, Envirex, and Euromatic) usually have a smaller ratio of disk to water surface area. The gas transfer through the exposed disk surface area is negligible when compared to the agitated surface area of the bulk water (Ouano, 1981). However, most of the other commercial applications of RDC in biological wastewater treatment are using much larger disk surface area. The water surface area becomes relatively insignificant.

Most authors assumed that the wetted disk surface area was equivalent to the effective contact area of the disks (Bintanja, 1974, Matsuo, 1982). The effective contact area is not necessarily equivalent to the wetted area in a gas-

liquid contact system. For some cases, e.g. in packed column, only the media surface covered by the fast-moving liquid film is considered as an effective contact area. the other parts of the wetted surface are considered as stagnant regions contributing very little to the absorption or desorption process. However, the disks used in this study had a relatively simpler surface configuration and the portions of the wetted disk surface were alternatively immersed into the bulk water by the disk rotation. Therefore, the equivalence between wetted area and effective contact area seems to be acceptable. Total contact area was then considered as the sum of free water surface area in the trough and the wetted area of the disks exposed to the air.

The disks used in this study were not flat disk, but disks with circular sockets of 3 cm in diameter and 1 cm in depth, which were unevenly distributed on the disk surface. The density of the sockets was higher toward the center of the disks. The area of each disk was determined individually by adding the extra area, provided by the sockets, to the area of a flat disk of the same diameter. The mean area of the disk was calculated by averaging the area of all the disks in the troughs. By visual observation, the wetted area of the disk showed a nearly circular water path. The wetted area of the discs was determined by directly measuring the radial width (H) of the wetted band

with a scale and converted into area by

$$A = m \quad (R^2 - \quad (R-H)^2)$$

where m = area of disk per area of a circle with the same radius, which is 1.47 for the disks used in this study. For practical reasons, only the outer disks were used for the measurement of the wetted band.

In a series of experiments at different rotational speeds, it was found that the wetted discs' area was independent of the rotational speed and was only a function of immersed depth. As shown in Table 5.5, the specific contact area, a (a ratio of effective contact area/bulk water volume), increases with decreasing submerged depth. Many other studies (Yamane, 1972; Bintanja, 1975; Zeevalkink, 1979) concluded the same observations as obtained in this study. The contribution of water surface increases as the immersed depth decreases. However, under normal operating conditions, Rh = 0.6 - 0.75, the free water surface contributes only 5 percent or less of the total contact area. In full-scale applications, the whole disk area is wetted, and the contribution of the water surface will be even more negligible (see Table 5.6). The surface turbulence increases the water surface area. Whereas there is no way to determine the actual area of a turbulent surface in the RDC unit, Rittmann (1983)

estimated that 15% of the K_La was attributed to the water surface transfer. Other authors assumed that the effect of water surface was insignificant (Matsuo, 1985, etc.). The extra water surface area provided by turbulence was not counted in the total contact area. The total contact area was therefore defined as the sum of the tank surface area and the total wetted area of discs that is exposed to the air. The specific contact area, a, was then calculated by dividing the values of wetted area by the volume of solution in the trough. As shown in Table 5.5, the specific contact area decreases with increasing immersed depth. However, due to the increasing number of sockets near the shaft, the specific contact area of Rh=0.68 is approximately equal to that of Rh=0.49.

Table 5.1 Data for Oxygen Transfer (Batch Study)

Water Volume (L)	Rotational Speed, w (rpm)	(rad/sec)	Immersed Depth (m)	Specific Area, a (1/m)	Rh i.e. (H/R)	(R-H)/R i.e. (1-Rh)	Water Temp. (C)	Room Temp. (C)	KLa 10-3 (1/sec)	KL 10-6 (m/sec)
5.0	2.46	0.258	0.11	245	0.59	0.41	23	23	0.30	1.22
5.0	11.00	1.152	0.11	245	0.59	0.41	22	22	3.29	13.43
5.0	12.80	1.340	0.11	245	0.59	0.41	22	22	4.90	20.00
5.0	13.60	1.424	0.11	245	0.59	0.41	23	23	4.96	20.24
5.0	14.99	1.570	0.11	245	0.59	0.41	27	23	6.87	28.04
5.0	16.80	1.759	0.11	245	0.59	0.41	21	21	7.57	30.90
5.0	17.25	1.806	0.11	245	0.59	0.41	21	21	7.79	31.80
4.0	0.23	0.024	0.09	290	0.49	0.51	30	25	0.32	1.11
4.0	0.36	0.038	0.09	290	0.49	0.51	25	25	0.33	1.14
4.0	0.78	0.082	0.09	290	0.49	0.51	30	25	0.50	1.72
4.0	0.81	0.085	0.09	290	0.49	0.51	20	23	0.53	1.83
4.0	1.22	0.128	0.09	290	0.49	0.51	30	25	0.60	2.07
4.0	2.45	0.257	0.09	290	0.49	0.51	20	23	0.75	2.59
4.0	3.48	0.364	0.09	290	0.49	0.51	20	20	1.45	5.00
4.0	4.13	0.433	0.09	290	0.49	0.51	30	25	1.72	5.93
4.0	5.27	0.552	0.09	290	0.49	0.51	20	23	2.32	8.00
4.0	5.70	0.597	0.09	290	0.49	0.51	21	21	2.53	8.72
4.0	5.90	0.618	0.09	290	0.49	0.51	20	23	2.46	8.48
4.0	6.58	0.689	0.09	290	0.49	0.51	20	20	3.05	10.52
4.0	7.68	0.804	0.09	290	0.49	0.51	25	25	3.75	12.93
4.0	7.90	0.827	0.09	290	0.49	0.51	20	25	3.87	13.33
4.0	9.10	0.953	0.09	290	0.49	0.51	20	24	4.60	15.86
4.0	9.93	1.040	0.09	290	0.49	0.51	20	25	5.62	19.38
4.0	11.17	1.170	0.09	290	0.49	0.51	22	22	5.91	20.38
4.0	11.94	1.250	0.09	290	0.49	0.51	21	21	6.57	22.66
4.0	13.66	1.430	0.09	290	0.49	0.51	20	20	8.80	30.34
4.0	15.25	1.597	0.09	290	0.49	0.51	30	25	9.70	33.45
4.0	15.64	1.638	0.09	290	0.49	0.51	25	25	10.38	35.78
4.0	17.19	1.800	0.09	290	0.49	0.51	20	20	11.28	38.90
3.0	0.27	0.028	0.07	343	0.39	0.61	25	25	0.33	0.96
3.0	0.45	0.047	0.07	343	0.39	0.61	20	25	0.47	1.37
3.0	0.82	0.085	0.07	343	0.39	0.61	20	25	0.59	1.72
3.0	1.64	0.172	0.07	343	0.39	0.61	25	25	0.93	2.71
3.0	2.44	0.256	0.07	343	0.39	0.61	20	25	1.18	3.44
3.0	4.89	0.512	0.07	343	0.39	0.61	20	23	2.39	6.95
3.0	6.77	0.709	0.07	343	0.39	0.61	20	25	3.69	10.76
3.0	11.75	1.230	0.07	343	0.39	0.61	20	20	6.62	19.29
3.0	12.60	1.319	0.07	343	0.39	0.61	22	22	7.50	21.85
3.0	14.10	1.477	0.07	343	0.39	0.61	20	24	9.30	27.11
3.0	14.40	1.508	0.07	343	0.39	0.61	22	22	9.16	26.71
3.0	15.95	1.670	0.07	343	0.39	0.61	20	25	10.00	29.15
3.0	16.50	1.728	0.07	343	0.39	0.61	19	19	11.30	32.94

Table 5.1 Data for Oxygen Transfer (Batch Study - continued)

Water Volume (L)	Rotational Speed, w (rpm)	(rad/sec)	Immersed Depth (m)	Specific Area, a (1/m)	Rh i.e. (H/R)	(R-H)/R i.e. (1-Rh)	Water Temp. (C)	Room Temp. (C)	KLa 10-3 (1/sec)	KL 10-6 (m/sec)
2.5	5.60	0.586	0.06	358	0.34	0.66	23	23	1.48	4.12
2.5	6.70	0.702	0.06	358	0.34	0.66	22	22	1.79	4.99
2.5	8.98	0.940	0.06	358	0.34	0.66	20	24	4.50	12.56
2.0	0.37	0.039	0.05	415	0.28	0.72	25	25	0.52	1.26
2.0	0.45	0.047	0.05	415	0.28	0.72	20	25	0.60	1.45
2.0	1.22	0.128	0.05	415	0.28	0.72	25	25	0.79	1.90
2.0	1.63	0.171	0.05	415	0.28	0.72	25	25	1.07	2.58
2.0	3.06	0.320	0.05	415	0.28	0.72	20	25	1.65	3.99
2.0	4.87	0.510	0.05	415	0.28	0.72	20	25	2.76	6.65
2.0	5.85	0.613	0.05	415	0.28	0.72	20	23	2.98	7.17
2.0	6.49	0.680	0.05	415	0.28	0.72	25	25	3.59	8.65
2.0	7.73	0.810	0.05	415	0.28	0.72	20	23	4.42	10.65
2.0	8.13	0.851	0.05	415	0.28	0.72	20	24	5.00	12.05
2.0	9.31	0.975	0.05	415	0.28	0.72	20	25	5.62	13.54
2.0	10.46	1.095	0.05	415	0.28	0.72	20	23	6.56	15.81
2.0	11.34	1.188	0.05	415	0.28	0.72	20	20	6.56	15.81
2.0	12.61	1.320	0.05	415	0.28	0.72	20	20	7.90	19.04
2.0	13.79	1.444	0.05	415	0.28	0.72	20	24	9.00	21.69
2.0	14.42	1.510	0.05	415	0.28	0.72	22	22	10.10	24.34
2.0	14.60	1.529	0.05	415	0.28	0.72	20	24	10.60	25.55
2.0	15.47	1.620	0.05	415	0.28	0.72	25	25	10.52	25.35
2.0	15.80	1.655	0.05	415	0.28	0.72	20	25	11.70	28.19
2.0	16.52	1.730	0.05	415	0.28	0.72	20	20	12.20	29.40
1.5	2.86	0.300	0.04	437	0.22	0.78	20	24	1.15	2.63
1.5	9.15	0.958	0.04	437	0.22	0.78	20	24	5.12	11.71
1.0	0.25	0.026	0.03	530	0.15	0.85	20	25	0.77	1.45
1.0	0.41	0.043	0.03	530	0.15	0.85	25	25	0.73	1.38
1.0	0.84	0.088	0.03	530	0.15	0.85	20	25	0.91	1.71
1.0	1.63	0.171	0.03	530	0.15	0.85	25	25	1.22	2.30
1.0	2.26	0.237	0.03	530	0.15	0.85	20	25	1.63	3.08
1.0	4.89	0.512	0.03	530	0.15	0.85	20	25	3.12	5.89
1.0	6.32	0.662	0.03	530	0.15	0.85	20	25	3.99	7.53
1.0	8.77	0.918	0.03	530	0.15	0.85	20	24	5.66	10.68
1.0	11.59	1.214	0.03	530	0.15	0.85	20	20	7.90	14.91
1.0	14.61	1.530	0.03	530	0.15	0.85	20	24	11.18	21.09
1.0	16.08	1.684	0.03	530	0.15	0.85	20	25	12.91	24.36

Table 5.2 Batch Study Data for VOCs Stripping

Rotational Speed, w (rpm)	(rad/s)	Immers. Depth (m)	Specif. Area, a (1/m)	Rh (H/R)	R1/R	Water Temp. (C)	Room Temp. (C)	KLa 10-3 (1/sec) TCE	PCE	EB	CLF	TCEN	KL 10-6 (1/sec) TCE	PCE	EB	CLF	TCEN
1.62	0.170	0.09	290	0.49	0.51	20	20	0.63	0.64	0.64			2.17	2.21	2.21		
3.50	0.367	0.09	290	0.49	0.51	20	20	0.87					2.99				
3.50	0.367	0.09	290	0.49	0.51	20	20	0.53		1.03			1.83		3.55		
4.49	0.470	0.09	290	0.49	0.51	20	21	1.26	1.23	0.84	0.95	0.86	4.34	4.24	2.90	3.28	2.97
7.76	0.813	0.09	290	0.49	0.51	20	22	2.33	2.25	2.10	2.01	1.90	8.03	7.76	7.24	6.93	6.55
10.80	1.131	0.09	290	0.49	0.51	20	20	3.30	3.35	2.92	2.90	3.00	11.38	11.55	10.07	10.00	10.34
11.22	1.175	0.09	290	0.49	0.51	20	20	3.99					13.76				
11.83	1.239	0.09	290	0.49	0.51	20	20	4.30	4.08	4.03	3.90	4.15	14.83	14.07	13.90	13.45	14.31
11.94	1.250	0.09	290	0.49	0.51	20	20	4.03	3.98	3.80	3.70	3.90	13.90	13.72	13.10	12.76	13.45
12.35	1.293	0.09	290	0.49	0.51	20	10	4.45					15.34				
12.80	1.340	0.09	290	0.49	0.51	20	20	4.60	4.60	4.30	4.10	4.30	15.86	15.86	14.83	14.14	14.83
13.75	1.440	0.09	290	0.49	0.51	20	20	4.92					16.97				
15.01	1.572	0.09	290	0.49	0.51	20	20	5.56	5.58	5.45			19.17	19.24	18.79		
15.24	1.596	0.09	290	0.49	0.51	20	20	5.80	5.90	5.70	5.60	5.77	20.00	20.34	19.66	19.31	19.90
16.03	1.679	0.09	290	0.49	0.51	20	20	6.50	6.40	6.10	5.90	6.32	22.41	22.07	21.03	20.34	21.79
16.42	1.720	0.09	290	0.49	0.51	20	20	7.10	6.80	6.40	6.30	6.45	24.48	23.45	22.07	21.72	22.24
16.49	1.727	0.09	290	0.49	0.51	20	20	6.10					21.03				
17.57	1.840	0.09	290	0.49	0.51	20	20	6.40					22.07				
17.86	1.870	0.09	290	0.49	0.51	20	20	6.60	6.20	6.60	6.20	6.60	22.76	21.38	22.76	21.38	22.76
17.99	1.884	0.09	290	0.49	0.51	20	20	8.70					30.00				
13.82	1.447	0.07	343	0.39	0.61	20	20	4.08	5.90	5.50			11.90	17.20	16.03		
13.37	1.400	0.06	378	0.34	0.66	20	20	5.30	4.70	5.60			14.02	12.43	14.81		
11.55	1.209	0.05	415	0.28	0.72	20	20	4.90	4.93	5.11			11.81	11.88	12.31		
12.49	1.308	0.05	415	0.28	0.72	20	20	6.10	6.50	6.03			14.70	15.66	14.53		
14.11	1.478	0.05	415	0.28	0.72	20	27	8.12	8.06	6.02		8.30	19.55	19.42	14.51		20.00
12.61	1.320	0.03	530	0.15	0.85	20	20	6.50	6.10	6.40			12.26	11.51	12.08		
5.83	0.611	0.03	530	0.15	0.85	20	26		3.00	2.90				5.66	5.47		
8.60	0.901	0.09	290	0.49	0.51	24	26	3.74	3.60	3.50	3.39	3.65	12.91	12.41	12.07	11.69	12.59
8.79	0.920	0.09	290	0.49	0.51	23	23	3.57	3.49	3.43	3.05	3.55	12.31	12.02	11.81	10.52	12.24
8.93	0.935	0.09	290	0.49	0.51	23	26	3.67	3.68	3.49		3.60	12.66	12.69	12.03		12.41
9.52	0.997	0.09	290	0.49	0.51	15	21	3.61	3.76	3.50	3.33	3.55	12.43	12.96	12.07	11.50	12.24
9.52	0.997	0.09	290	0.49	0.51	21	21	3.97	4.22	3.93	3.60	3.90	13.69	14.54	13.55	12.40	13.43
9.52	0.997	0.09	290	0.49	0.51	27	21	4.06	4.20	4.03	3.81	3.89	13.99	14.49	13.88	13.14	13.42
11.08	1.160	0.09	290	0.49	0.51	22	22	4.35	4.42	4.30			15.00	15.24	14.83		
11.22	1.175	0.09	290	0.49	0.51	20	20	3.99					13.76				
11.65	1.220	0.09	290	0.49	0.51	10	20	3.70	4.00	3.80			12.76	13.79	13.10		
11.83	1.239	0.09	290	0.49	0.51	20	20	4.30	4.18	4.03			14.83	14.41	13.90		
11.94	1.250	0.09	290	0.49	0.51	20	20	4.03	3.98	3.80			13.90	13.72	13.10		
13.75	1.440	0.09	290	0.49	0.51	20	20	4.92					16.97				
14.02	1.468	0.09	290	0.49	0.51	17	17	4.90	4.20	4.50			16.90	14.48	15.52		

Table 5.3 A Comparison of the Results of the Log Deficit Method and Exponential Method

	Log Deficit Method		Exponential Method	
w, (rpm)	Cs (ug/l)	n	Cs (ug/l)	n
Trichloroethylene				
4.49	0.1	1.26	0	1.26
7.76	1.4	2.34	0	2.33
12.80	0.9	4.65	0	4.60
16.42	1.7	7.01	0	7.10
17.86	1.2	6.68	0	6.60
Chloroform				
4.49	1.0	1.26	0	1.26
7.76	2.1	2.11	0	2.01
12.80	2.8	4.21	0	4.10
16.42	1.7	6.35	0	6.30
17.86	4.3	6.27	0	6.20

Table 5.4 Removal of VOCs in Continuous Flow 4-Stage RBC System

No. of Run	12-3			11-29			11-28			11-27-1			11-27-2			11-26		
w,rpm	12.4			0.82			4.59			0			13.2			1.17		
rad/s	1.3			0.085			0.48			0			1.38			0.12		
Flow,ml/min	386.4			363			380			382			358			382		
Room T, C	17			20			20			20			20			21		
Tank T, C	17			21			16			20			20			21		
Basin T, C	14			15			15			16			18			17		
1st Stage	Rem	KLa 1/sec 10-3	KL m/sec 10-6	Rem	KLa 1/sec 10-3	KL m/sec 10-6	Rem	KLa 1/sec 10-3	KL m/sec 10-6	Rem	KLa 1/sec 10-3	KL m/sec 10-6	Rem	KLa 1/sec 10-3	KL m/sec 10-6	Rem	KLa 1/sec 10-3	KL m/sec 10-6
Volum,L	6			6			6			6			6.2			6		
CLF	0.620	4.876	16.87	0.215	0.276	0.96	0.541	1.244	4.30	0.000	0.000	0.00	0.837	4.942	17.10	0.271	0.394	1.36
TCEH	0.835	5.424	18.77	0.233	0.306	1.06	0.545	1.264	4.37	0.000	0.000	0.00	0.848	5.369	18.58	0.297	0.448	1.55
TCE	0.834	5.400	18.69	0.297	0.426	1.47	0.573	1.416	4.90	0.000	0.000	0.00	0.850	5.453	18.87	0.356	0.587	2.03
PCE	0.836	5.475	18.95	0.344	0.529	1.83	0.589	1.513	5.23	0.012	0.013	0.04	0.853	5.584	19.32	0.403	0.716	2.48
EB	0.828	5.149	17.82	0.331	0.499	1.73	0.584	1.482	5.13	0.018	0.019	0.07	0.844	5.207	18.02	0.386	0.667	2.31
2nd Stage																		
Volum,L	6			6			6			6			6			6		
CLF	0.817	4.781	16.54	0.190	0.237	0.82	0.483	0.986	3.41	0.014	0.015	0.05	0.845	5.226	18.08	0.230	0.317	1.10
TCEH	0.831	5.290	18.31	0.211	0.270	0.93	0.493	1.026	3.55	0.022	0.024	0.08	0.855	5.663	19.60	0.259	0.371	1.28
TCE	0.832	5.321	18.41	0.271	0.374	1.29	0.529	1.186	4.10	0.042	0.047	0.16	0.851	5.486	18.98	0.337	0.539	1.87
PCE	0.835	5.425	18.77	0.311	0.455	1.57	0.522	1.153	3.99	0.097	0.114	0.39	0.857	5.756	19.92	0.378	0.645	2.23
EB	0.826	5.096	17.63	0.302	0.437	1.51	0.52	1.144	3.96	0.099	0.117	0.40	0.851	5.486	18.98	0.363	0.605	2.09
3rd Stage																		
Volum,L	6			6			6			6			6			6		
CLF	0.815	4.735	16.38	0.190	0.237	0.82	0.503	1.068	3.70	0.019	0.021	0.07	0.847	5.328	18.43	0.262	0.377	1.30
TCEH	0.830	5.226	18.08	0.210	0.268	0.93	0.513	1.112	3.85	0.023	0.025	0.09	0.857	5.767	19.96	0.289	0.431	1.49
TCE	0.831	5.282	18.28	0.264	0.362	1.25	0.519	1.139	3.94	0.026	0.028	0.10	0.851	5.496	19.02	0.337	0.539	1.87
PCE	0.834	5.400	18.69	0.309	0.451	1.56	0.520	1.144	3.96	0.040	0.044	0.15	0.858	5.815	20.12	0.378	0.645	2.23
EB	0.825	5.071	17.55	0.303	0.438	1.52	0.518	1.134	3.93	0.039	0.043	0.15	0.853	5.584	19.32	0.367	0.615	2.13
4th Stage																		
Volum,L	6			6			6			6						6		
CLF	NA	NA	NA	0.260	0.252	0.87	0.493	1.026	3.55	NA	NA	NA	NA	NA	NA	0.202	0.269	0.93
TCEH	NA	NA	NA	0.228	0.298	1.03	0.502	1.064	3.68	NA	NA	NA	NA	NA	NA	0.225	0.308	1.07
TCE	NA	NA	NA	0.291	0.414	1.43	0.522	1.153	3.99	NA	NA	NA	NA	NA	NA	0.285	0.423	1.46
PCE	NA	NA	NA	0.337	0.513	1.77	0.525	1.167	4.04	NA	NA	NA	NA	NA	NA	0.327	0.516	1.78
EB	NA	NA	NA	0.328	0.492	1.70	0.515	1.121	3.88	NA	NA	NA	NA	NA	NA	0.317	0.492	1.70

* NA: the residuals were lower than detection limits

Table 5.4 Reanval of VOCs in Continuous Flow 4-Stage RDC System (continued)

No. of Run	11-25			11-19			11-12			11-11			11-3			11-10		
w,rpm	3.99			7.37			5.52			5.58			4.89			1.63		
rad/s	0.42			0.77			0.58			0.59			0.51			0.17		
Flow,ml/min	382			382.2			400			237			131.8			132.2		
Room T, C	20			15			20			17			20			21		
Tank T, C	20			16			16			23			19			20		
Basin T, C	15			14			15			16			15			14		
1st Stage	Rem	KLa 1/sec 10-3	KL m/sec 10-6	Rem	KLa 1/sec 10-3	KL m/sec 10-6	Rem	KLa 1/sec 10-3	KL m/sec 10-6	Rem	KLa 1/sec 10-3	KL m/sec 10-6	Rem	KLa 1/sec 10-3	KL m/sec 10-6	Rem	KLa 1/sec 10-3	KL m/sec 10-6
Volum,L	6			6			6			5.9			6			5.9		
CLF	0.484	0.995	3.44	0.714	2.644	9.15	0.603	1.688	5.84	0.715	1.680	5.81	0.817	1.639	5.67	0.646	0.691	2.36
TCEN	0.488	1.011	3.50	0.726	2.813	9.73	0.601	1.674	5.79	0.709	1.631	5.64	0.815	1.613	5.58	0.629	0.633	2.19
TCE	0.502	1.070	3.70	0.732	2.900	10.03	0.620	1.813	6.27	0.739	1.896	6.56	0.827	1.750	6.06	0.675	0.776	2.68
PCE	0.492	1.028	3.56	0.735	2.943	10.18	0.631	1.900	6.57	0.753	2.041	7.06	0.831	1.800	6.23	0.717	0.946	3.27
EB	0.488	1.011	3.50	0.725	2.799	9.68	0.622	1.828	6.33	0.742	1.925	6.66	0.827	1.750	6.06	0.707	0.901	3.12
2nd Stage																		
Volum,L	6			6			6			5.75			5.96			5.75		
CLF	0.447	0.856	2.96	0.640	1.887	6.53	0.557	1.395	4.83	0.648	1.232	4.26	0.742	1.053	3.64	0.633	0.644	2.23
TCEN	0.454	0.883	3.06	0.662	2.079	7.20	0.567	1.457	5.04	0.671	1.365	4.72	0.750	1.098	3.80	0.653	0.703	2.43
TCE	0.461	0.907	3.14	0.663	2.085	7.21	0.579	1.529	5.29	0.674	1.384	4.79	0.760	1.159	4.01	0.690	0.831	2.88
PCE	0.462	0.911	3.15	0.668	2.136	7.39	0.585	1.569	5.43	0.676	1.397	4.83	0.762	1.172	4.06	0.716	0.942	3.26
EB	0.456	0.889	3.07	0.654	2.007	6.94	0.577	1.517	5.25	0.663	1.317	4.56	0.757	1.141	3.95	0.706	0.897	3.10
3rd Stage																		
Volum,L	6			6			6			5.6			5.65			5.65		
CLF	0.434	0.814	2.82	0.648	1.954	6.76	0.545	1.331	4.61	0.650	1.243	4.30	0.7456	1.073	3.71	0.677	0.793	2.71
TCEN	0.443	0.844	2.92	0.677	2.225	7.70	0.559	1.408	4.87	0.664	1.323	4.58	0.7566	1.138	3.94	0.690	0.831	2.88
TCE	0.447	0.858	2.97	0.676	2.215	7.66	0.569	1.467	5.08	0.683	1.442	4.99	0.7599	1.159	4.01	0.715	0.937	3.24
PCE	0.452	0.875	3.03	0.678	2.235	7.74	0.574	1.497	5.18	0.686	1.463	5.06	0.7603	1.161	4.02	0.725	0.985	3.41
EB	0.445	0.851	2.94	0.664	2.098	7.26	0.566	1.449	5.01	0.679	1.416	4.90	0.7577	1.145	3.96	0.718	0.951	3.29
4th Stage																		
Volum,L	6			6			6			5.6						5.65		
CLF	0.421	0.772	2.67	0.638	1.871	6.47	0.525	1.228	4.25	0.675	1.390	4.81	NA	NA	NA	0.641	0.667	2.31
TCEN	0.421	0.772	2.67	0.671	2.165	7.49	0.539	1.299	4.50	0.701	1.570	5.43	NA	NA	NA	0.667	0.748	2.59
TCE	0.436	0.820	2.84	0.666	2.117	7.33	0.546	1.336	4.62	0.700	1.562	5.41	NA	NA	NA	0.712	0.923	3.19
PCE	0.439	0.830	2.87	0.669	2.146	7.42	0.549	1.353	4.68	0.694	1.518	5.25	NA	NA	NA	0.744	1.085	3.76
EB	0.433	0.810	2.80	0.663	2.089	7.23	0.541	1.310	4.53	0.685	1.456	5.04	NA	NA	NA	0.749	1.114	3.86

* NA: the resid

Table 5.4 Removal of VOCs in Continuous Flow 4-Stage RDC System (continued)

No. of Run	10-25			10-23			10-18			10-15			10-11			10-8		
w,rpm	1.63			1.63			12.25			11.7			11.15			7.57		
rad/s	0.17			0.17			1.28			1.23			1.17			0.79		
Flow,ml/min	324.1			249.6			224			203			328.57			130.5		
Room T, C	22			15			25			20			21			20		
Tank T, C	20			15			20			20			20			19		
Basin T, C	15			13			15			15			15			15		
1st Stage	Rem	KLa 1/sec 10-3	KL m/sec 10-6	Rem	KLa 1/sec 10-3	KL m/sec 10-6	Rem	KLa 1/sec 10-3	KL m/sec 10-6	Rem	KLa 1/sec 10-3	KL m/sec 10-6	Rem	KLa 1/sec 10-3	KL m/sec 10-6	Rem	KLa 1/sec 10-3	KL m/sec 10-6
Volum,L	6			5.9			6.05			5.77			5.8			5.74		
CLF	0.324	0.431	1.49	0.384	0.404	1.40	0.885	4.726	16.35	0.896	5.052	17.48	0.833	4.710	16.30	0.886	2.945	10.19
TCEN	0.343	0.470	1.63	0.422	0.515	1.78	0.893	5.172	17.90	0.907	5.719	19.79	0.8417	5.020	17.37	0.888	3.004	10.40
TCE	0.404	0.610	2.11	0.478	0.646	2.23	0.899	5.481	18.96	0.910	5.914	20.46	0.8499	5.346	18.50	0.892	3.130	10.83
PCE	0.457	0.758	2.62	0.558	0.890	3.08	0.907	6.040	20.90	0.916	6.411	22.18	0.8577	5.691	19.69	0.893	3.162	10.94
EB	0.445	0.722	2.50	0.563	0.908	3.14	0.900	5.554	19.22	0.910	5.900	20.42	0.8485	5.288	18.30	0.890	3.066	10.61
2nd Stage																		
Volum,L	6.3			5.9			5.86			6.1			6.1			5.93		
CLF	0.5097	0.936	3.24	0.413	0.496	1.72	0.837	3.169	10.96	0.815	2.583	8.94	0.698	2.182	7.55	0.842	2.019	6.99
TCEN	0.4795	0.829	2.87	0.433	0.538	1.86	0.858	3.729	12.90	0.853	3.403	11.77	0.718	2.404	8.32	0.858	2.290	7.92
TCE	0.5606	1.149	3.97	0.485	0.664	2.30	0.858	3.729	12.90	0.845	3.197	11.06	0.718	2.404	8.32	0.859	2.308	7.99
PCE	0.5837	1.262	4.37	0.521	0.767	2.65	0.860	3.791	13.12	0.851	3.349	11.59	0.718	2.404	8.32	0.861	2.347	8.12
EB	0.5718	1.202	4.16	0.502	0.711	2.46	0.847	3.416	11.82	0.835	2.967	10.27	0.704	2.246	7.77	0.874	2.628	9.09
3rd Stage																		
Volum,L	5.6			5.8			5.8			5.8			6			5.8		
CLF	0.4846	0.846	2.93	0.473	0.633	2.19	0.872	4.200	14.53	0.866	3.790	13.11	0.792	3.595	12.44	0.848	2.114	7.31
TCEN	0.4934	0.877	3.03	0.497	0.697	2.41	0.889	4.932	17.07	0.894	4.945	17.11	0.81	4.025	13.93	0.866	2.449	8.47
TCE	0.527	1.003	3.47	0.541	0.831	2.88	0.882	4.612	15.96	0.887	4.603	15.93	0.808	3.973	13.75	0.858	2.290	7.92
PCE	0.5394	1.054	3.65	0.575	0.954	3.30	0.890	4.993	17.28	0.890	4.744	16.42	0.807	3.948	13.66	0.861	2.347	8.12
EB	0.5302	1.016	3.52	0.561	0.901	3.12	0.872	4.219	14.60	0.868	3.856	13.34	0.801	3.800	13.15	0.910	3.831	13.26
4th Stage																		
Volum,L	5.6			5.7									5.9					
CLF	0.466	0.786	2.72	0.319	0.330	1.14	NA	NA	NA	NA	NA	NA	0.76	2.990	10.35	NA	NA	NA
TCEN	0.476	0.818	2.83	0.318	0.329	1.14	NA	NA	NA	NA	NA	NA	0.786	3.468	12.00	NA	NA	NA
TCE	0.518	0.968	3.35	0.535	0.811	2.81	NA	NA	NA	NA	NA	NA	0.771	3.179	11.00	NA	NA	NA
PCE	0.535	1.036	3.58	0.379	0.430	1.49	NA	NA	NA	NA	NA	NA	NA	NA	NA	NA	NA	NA
EB	0.523	0.987	3.42	0.371	0.416	1.44	NA	NA	NA	NA	NA	NA	0.797	3.707	12.83	NA	NA	NA

* NA: the resid

Table 5.4 Removal of VOCs in Continuous Flow 4-Stage RBC System (continued)

No. of Run	10-6			9-21			9-19			9-18			9-15			9-14		
w,rpm	12.04			8.08			9.12			9.78			8.28			11.14		
rad/s	1.26			0.92			0.96			1.02			0.87			1.17		
Flow,ml/min	131			206			210			373.5			252			327		
Room T, C	20			20			23			18			21			19		
Tank T, C	20			20			23			18			21			19		
Basin T, C	16			15			18			18			15			15		
1st Stage	Rem	KLa 1/sec 10-3	KL m/sec 10-6	Rem	KLa 1/sec 10-3	KL m/sec 10-6	Rem	KLa 1/sec 10-3	KL m/sec 10-6	Rem	KLa 1/sec 10-3	KL m/sec 10-6	Rem	KLa 1/sec 10-3	KL m/sec 10-6	Rem	KLa 1/sec 10-3	KL m/sec 10-6
Volum,L	5.8			5.8			6			6			5.9			6		
CLF	0.924	4.544	15.72	0.840	3.108	10.75	0.854	3.412	11.81	0.782	3.730	12.91	0.815	3.136	10.85	0.793	3.480	12.04
TCEN	0.928	4.816	16.66	0.850	3.354	11.61	0.864	3.706	12.82	0.801	4.176	14.45	0.820	3.243	11.22	0.799	3.611	12.49
TCE	0.928	4.891	16.89	0.852	3.408	11.79	0.872	3.974	13.75	0.800	4.150	14.36	0.833	3.551	12.29	0.805	3.750	12.98
PCE	0.929	4.911	16.99	0.856	3.519	12.18	0.880	4.278	14.80	0.818	4.663	16.14	0.849	4.002	13.85	0.806	3.774	13.06
EB	0.925	4.629	16.02	0.850	3.354	11.61	0.873	4.010	13.87	0.809	4.403	15.24	0.852	4.099	14.18	0.801	3.656	12.65
2nd Stage																		
Volum,L	5.8			6			6			6			5.8			6		
CLF	0.906	3.628	12.55	0.816	2.622	9.07	0.802	2.368	8.20	0.745	3.031	10.49	0.671	1.452	5.02	0.749	2.711	9.38
TCEN	0.919	4.271	14.78	0.834	2.978	10.30	0.840	3.071	10.63	0.775	3.574	12.37	0.675	1.478	5.12	0.751	2.740	9.48
TCE	0.926	4.711	16.30	0.832	2.932	10.14	0.842	3.100	10.73	0.773	3.533	12.22	0.681	1.520	5.26	0.760	2.876	9.95
PCE	0.923	4.512	15.61	0.833	2.946	10.19	0.840	3.073	10.63	0.782	3.722	12.88	0.680	1.513	5.23	0.750	2.725	9.43
EB	0.918	4.214	14.58	0.820	2.697	9.33	0.840	3.057	10.58	0.768	3.434	11.88	0.675	1.478	5.12	0.754	2.784	9.63
3rd Stage																		
Volum,L				5.7			5.9			6			5.7			6		
CLF	NA	NA	NA	0.811	2.540	8.79	0.795	2.262	7.83	0.741	2.968	10.27	0.764	2.305	7.97	0.782	3.258	11.27
TCEN	NA	NA	NA	0.831	2.911	10.07	0.837	2.995	10.36	0.770	3.473	12.02	0.776	2.466	8.53	0.797	3.566	12.34
TCE	NA	NA	NA	0.828	2.850	9.86	0.837	3.002	10.39	0.758	3.250	11.24	0.779	2.509	8.68	0.796	3.544	12.26
PCE	NA	NA	NA	0.828	2.850	9.86	0.835	2.948	10.20	0.770	3.473	12.02	0.770	2.383	8.25	0.799	3.397	11.75
EB	NA	NA	NA	0.814	2.591	8.96	0.835	2.952	10.21	0.756	3.215	11.12	0.763	2.292	7.93	0.780	3.220	11.14
4th Stage																		
Volum,L							5.9			6			5.6			6		
CLF	NA	NA	NA	NA	NA	NA	0.827	2.789	9.65	0.725	2.735	9.46	0.746	2.091	7.23	0.746	2.668	9.23
TCEN	NA	NA	NA	NA	NA	NA	0.845	3.180	11.00	0.749	3.096	10.71	0.771	2.397	8.29	0.777	3.165	10.95
TCE	NA	NA	NA	NA	NA	NA	0.829	2.828	9.79	0.674	2.145	7.42	0.765	2.317	8.02	0.768	3.007	10.40
PCE	NA	NA	NA	NA	NA	NA	NA	NA	NA	0.679	2.195	7.59	0.746	2.091	7.23	0.755	2.799	9.69
EB	NA	NA	NA	NA	NA	NA	NA	NA	NA	0.638	1.829	6.33	0.743	2.058	7.12	0.747	2.682	9.28

* NA: the resid

Table 5.5 Data of Pilot Plant at Different Depth of Immersion

Immersed Depth, H(m)	0.03	0.04	0.05	0.06	0.07	0.08	0.09	0.11	0.13
Water Volume, V(L)	1.00	1.51	2.00	2.51	3.00	3.50	4.00	5.01	6.00
Immersed Depth/Radius, Rh	0.15	0.22	0.28	0.34	0.39	0.44	0.49	0.59	0.68
Total Contact Area, A(m2)	0.52	0.70	0.84	0.95	1.03	1.10	1.16	1.45	1.73
Disk Area/Total Area	0.93	0.94	0.95	0.95	0.95	0.95	0.95	0.95	0.95
Surface Area/Total Area	0.07	0.06	0.05	0.05	0.05	0.05	0.05	0.05	0.05
Disc Area/Surface Area	13	16	18	19	20	20	21	21	21
Specific Contact Area, a(1/m)	523	464	418	377	344	315	289	290	288

* "a" was calculated with total contact area

Table 5.6 Data of Full-Scale RDC for Different Immersed Depth

immerseddepth, H (m)	1.08	0.91	0.71	0.51	0.28
contactarea, (m^2)	12633	13528	14455	15329	16156
of which contributed by:					
the discs (%)	99.7	99.8	99.8	99.9	99.9
the surface of the trough(%)	0.3	0.2	0.2	0.1	0.1
water volume, V (m^2)	23.0	18.7	13.7	9.0	4.3
specific surface, a (1/m)	549	723	1055	1710	3748
immersed depth/radius, Rh	0.59	0.49	0.39	0.28	0.15

*calculation were based on the high density RDC unit of Lyco, Inc.

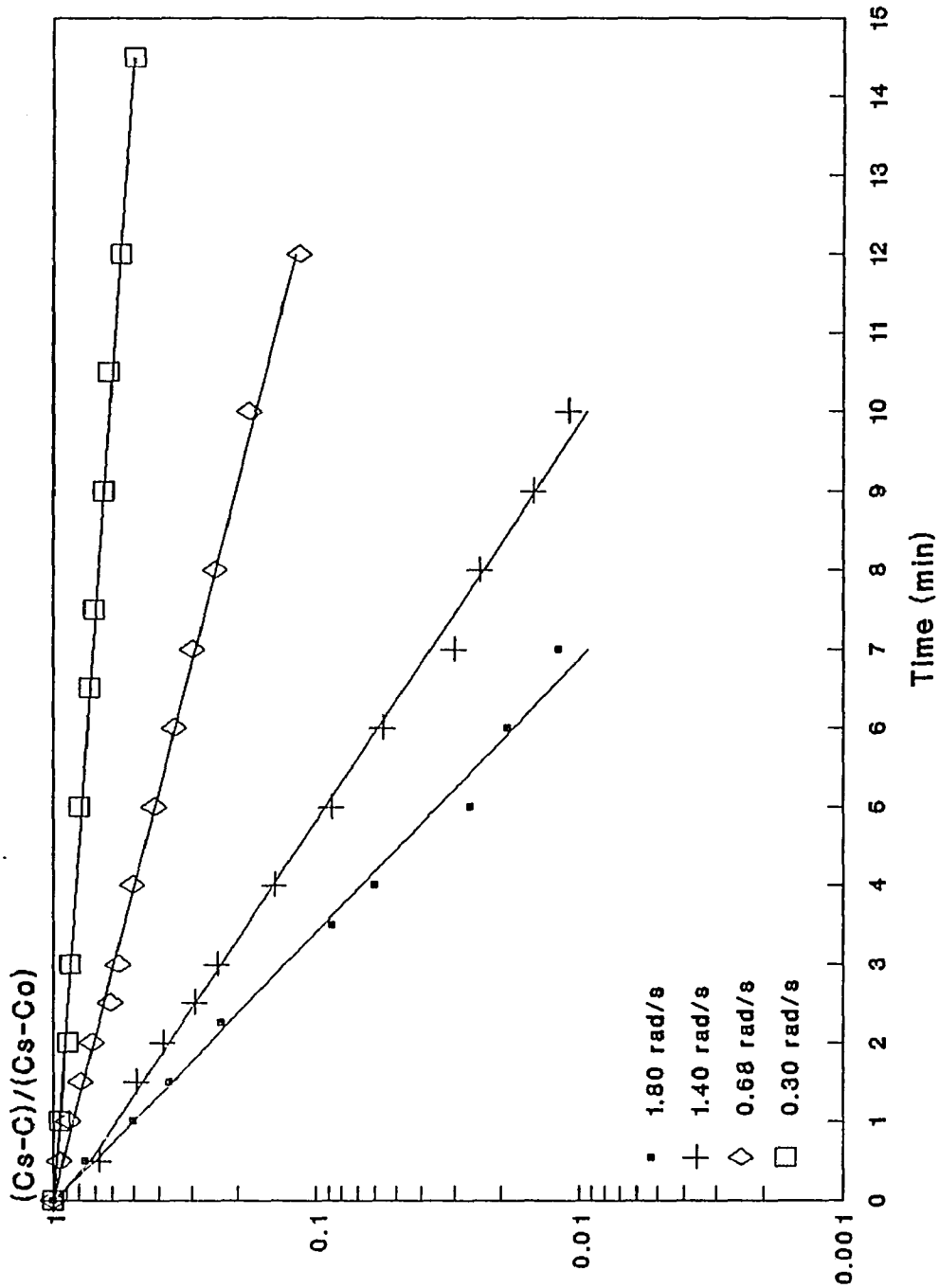


Fig 5.1 KLa determination for Oxygen
(20 C, 4L)
(Cs-C)/(Cs-Co)
1
0.1
0.01
0.001
0
1
2
3
4
5
6
7
8
9
10
11
12
13
14
15
Time (min)
1.80 rad/s
1.40 rad/s
0.68 rad/s
0.30 rad/s

Fig 5.2 KLa Determination for TCE
(20 C, Rh = 0.49)

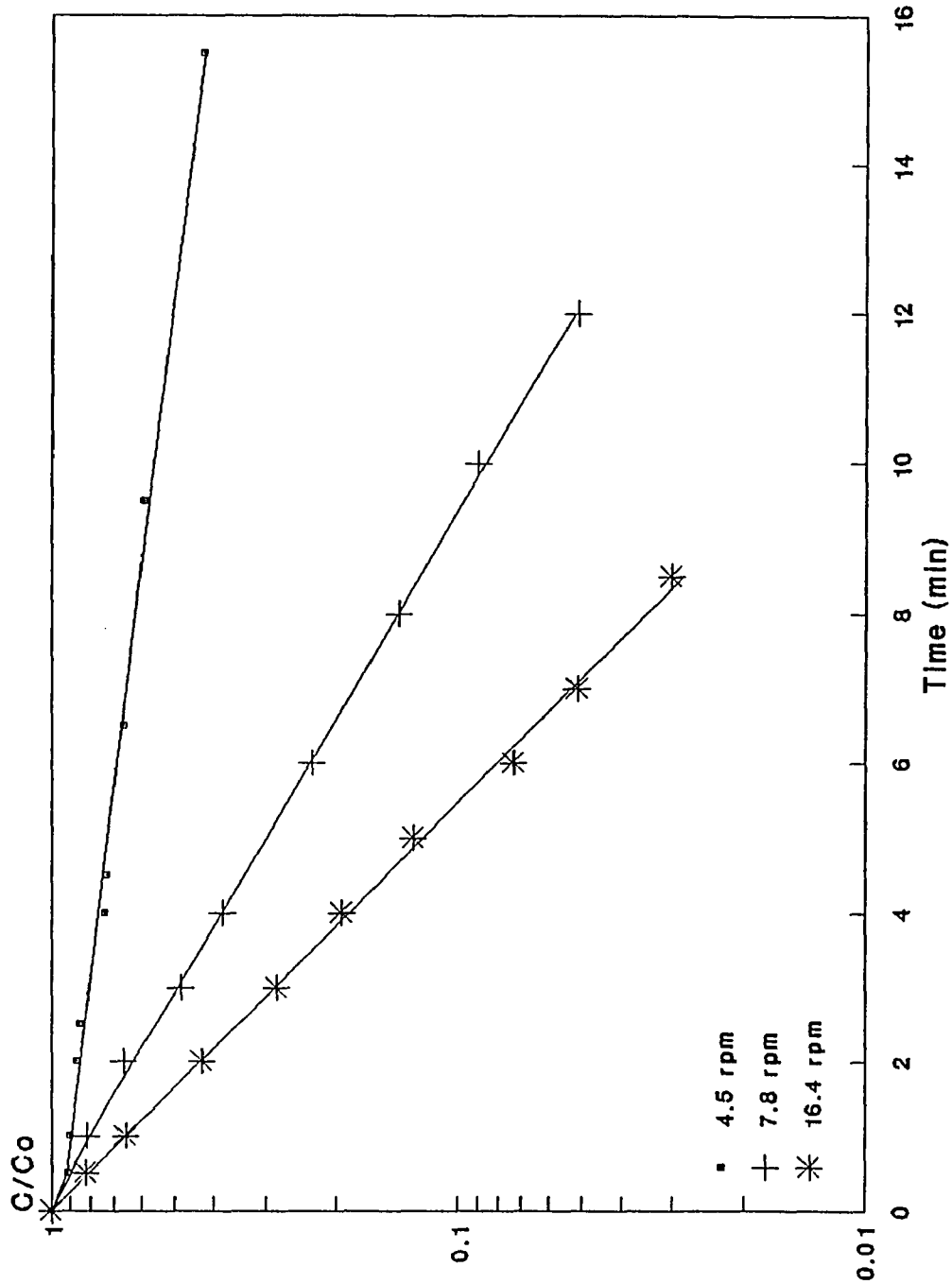


Fig 5.3 KLa Determination for PCE
(20 C, Rh = 0.49)

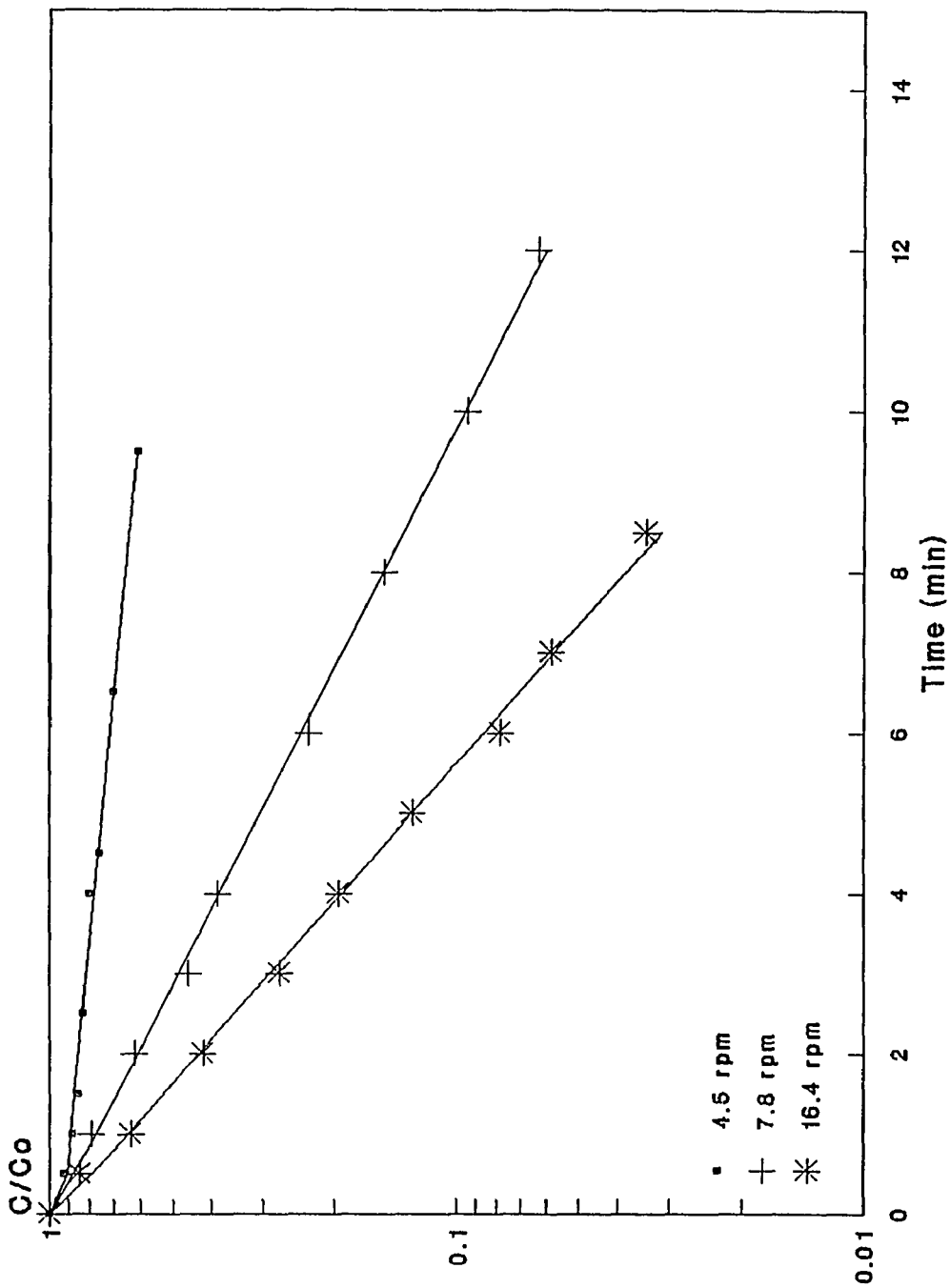


Fig 5.4 KLa Determination for EB
(20 C, Rh = 0.49)

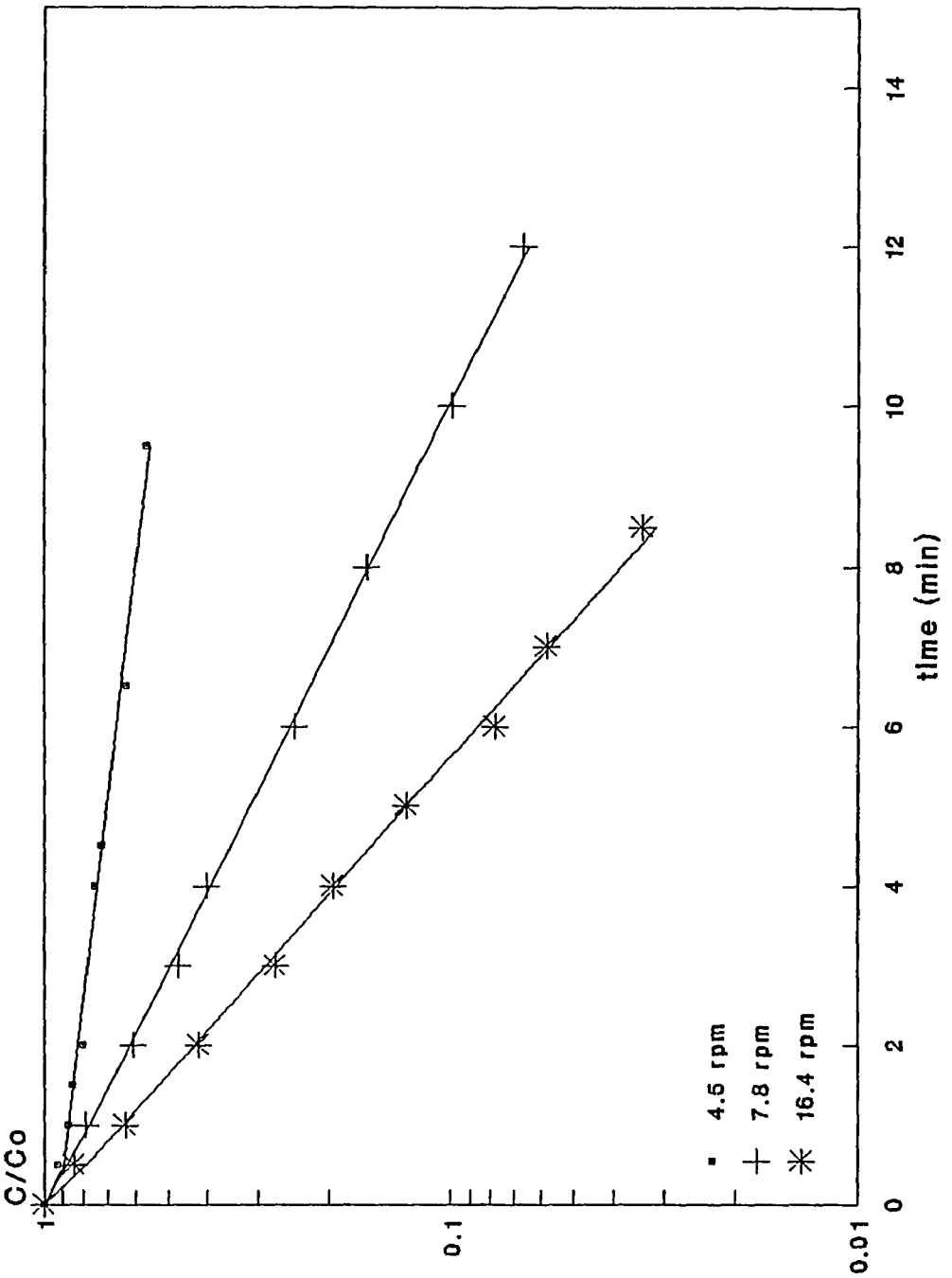


Fig 5.5 KLa for TCE, PCE, EB, TCEN, CLF
(21 C, 9.52 rpm, Rh = 0.49)

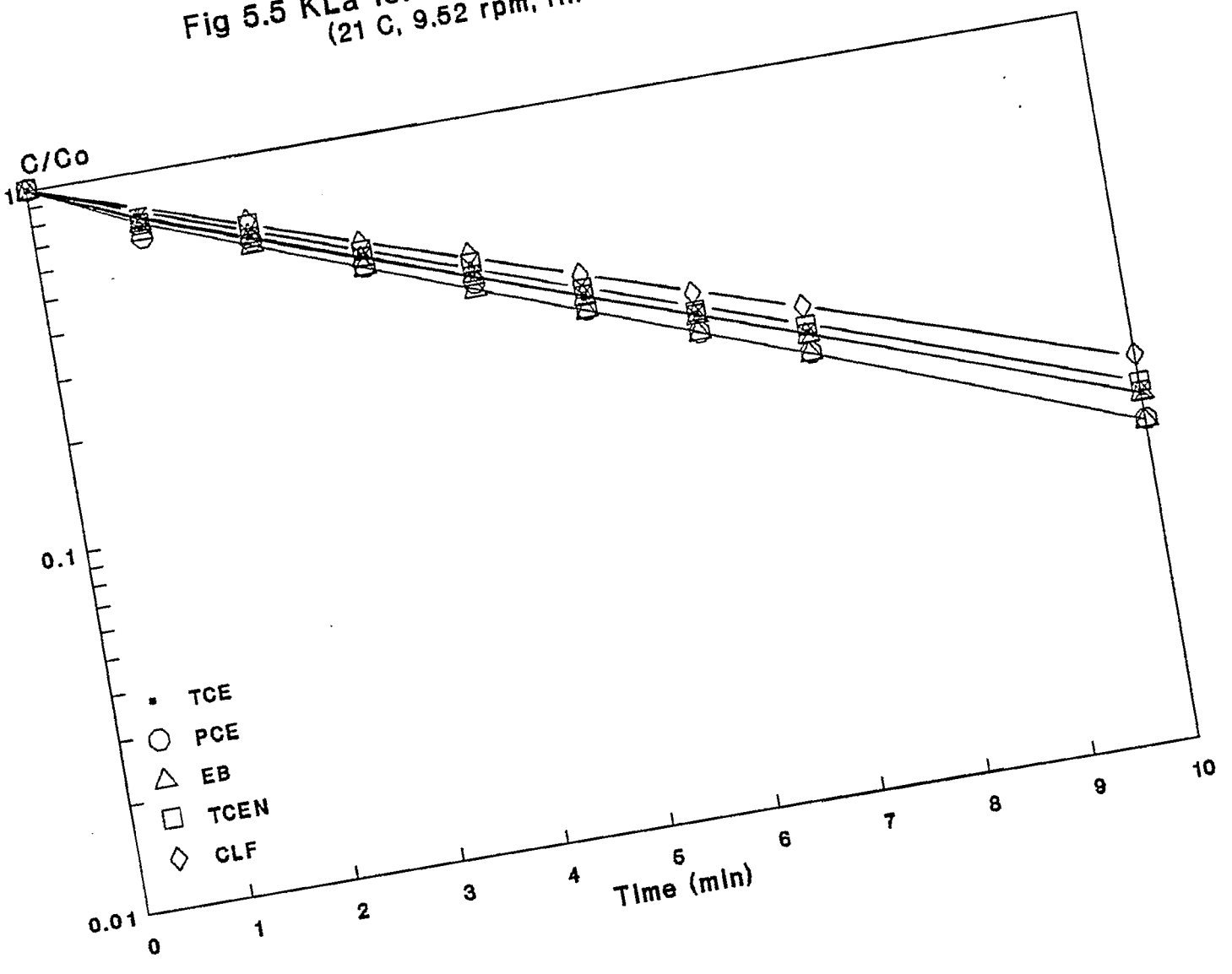


Fig 5.6 KLa Determination by Direct Method (for Oxygen)

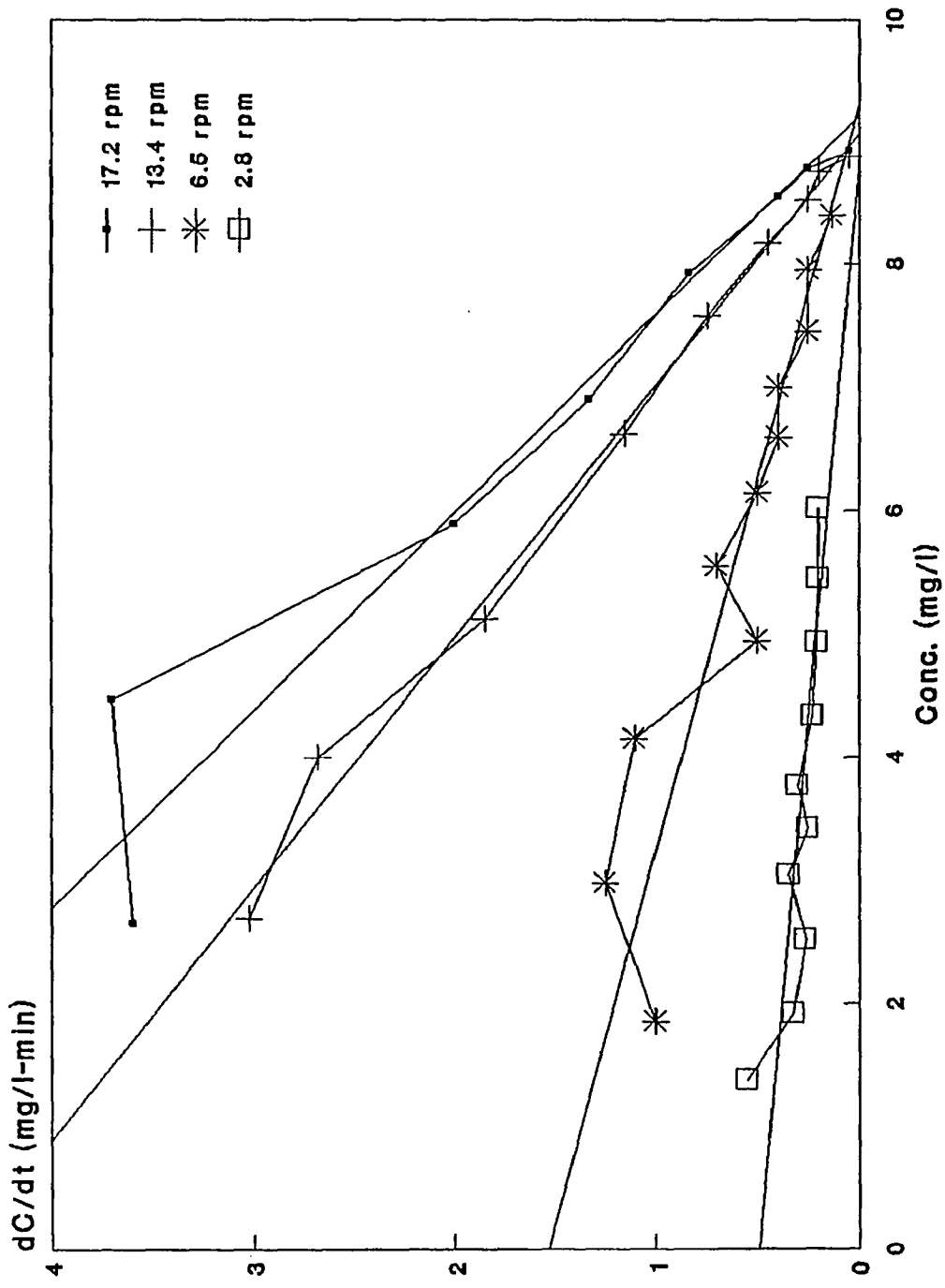


Fig 5.7 Response Curve of Batch Test
(oxygen, Rh=0.49, 20 C)

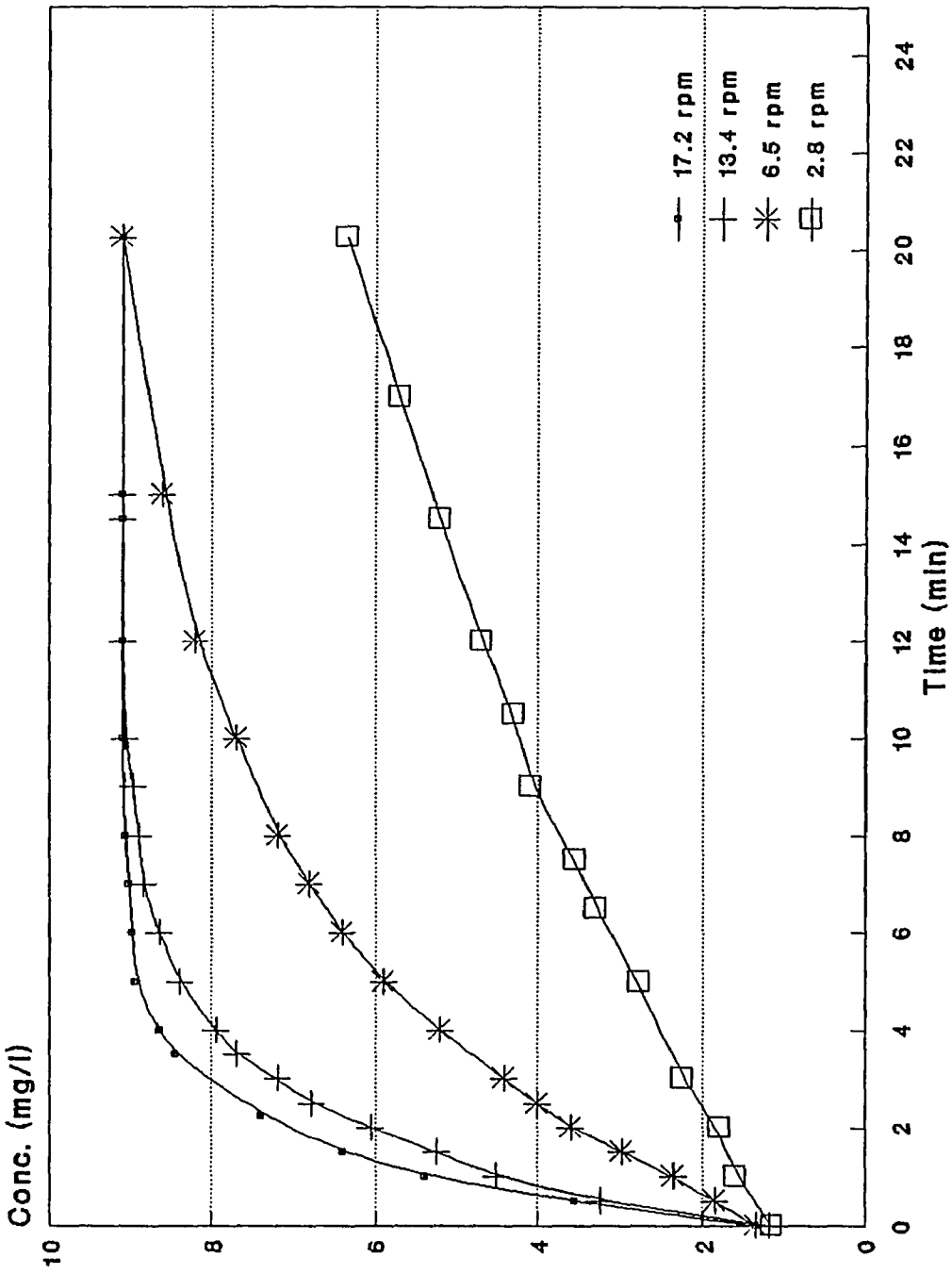


Fig 5.8 Step Response Curve of RDC
(Tracer Study)

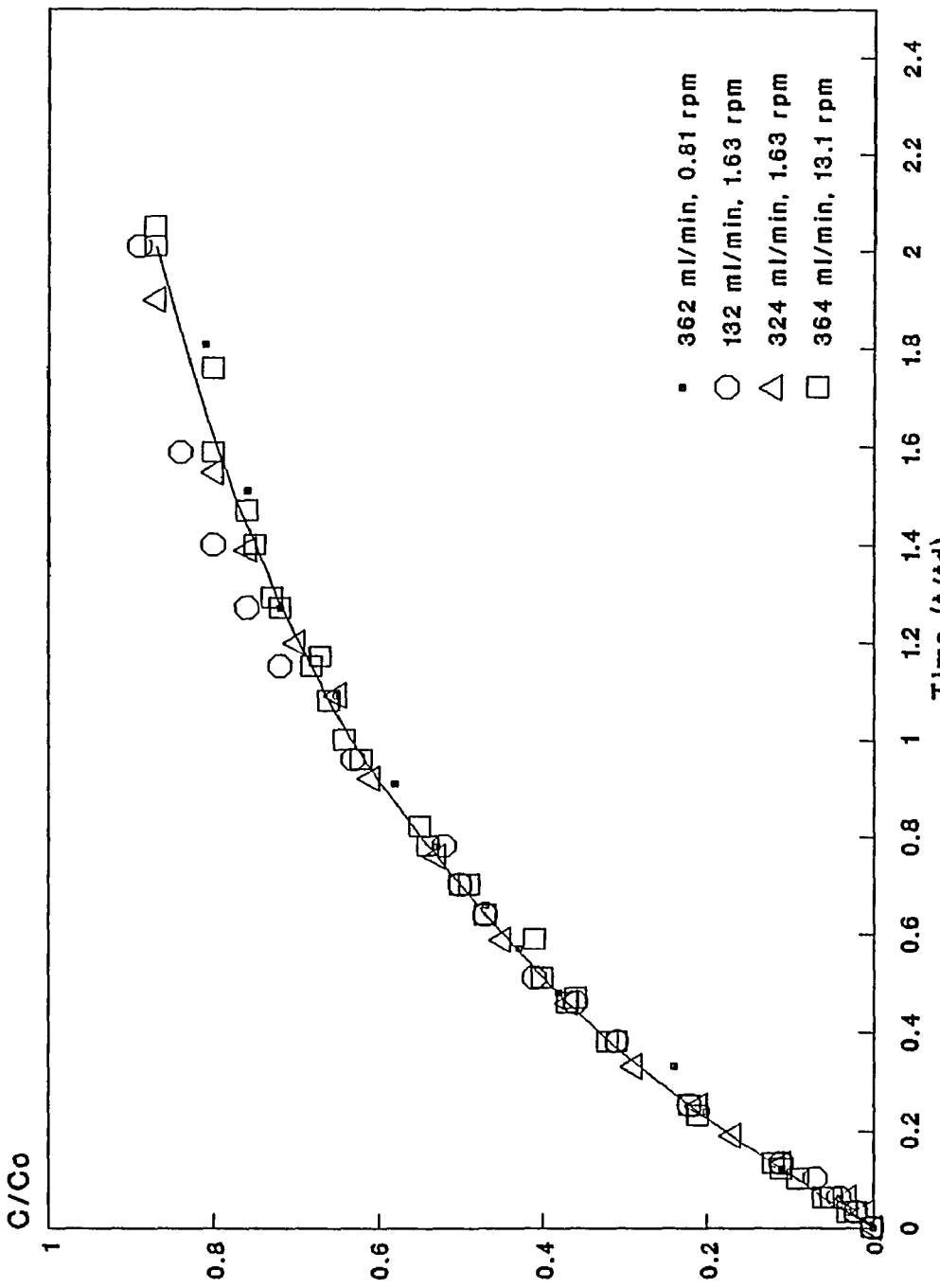
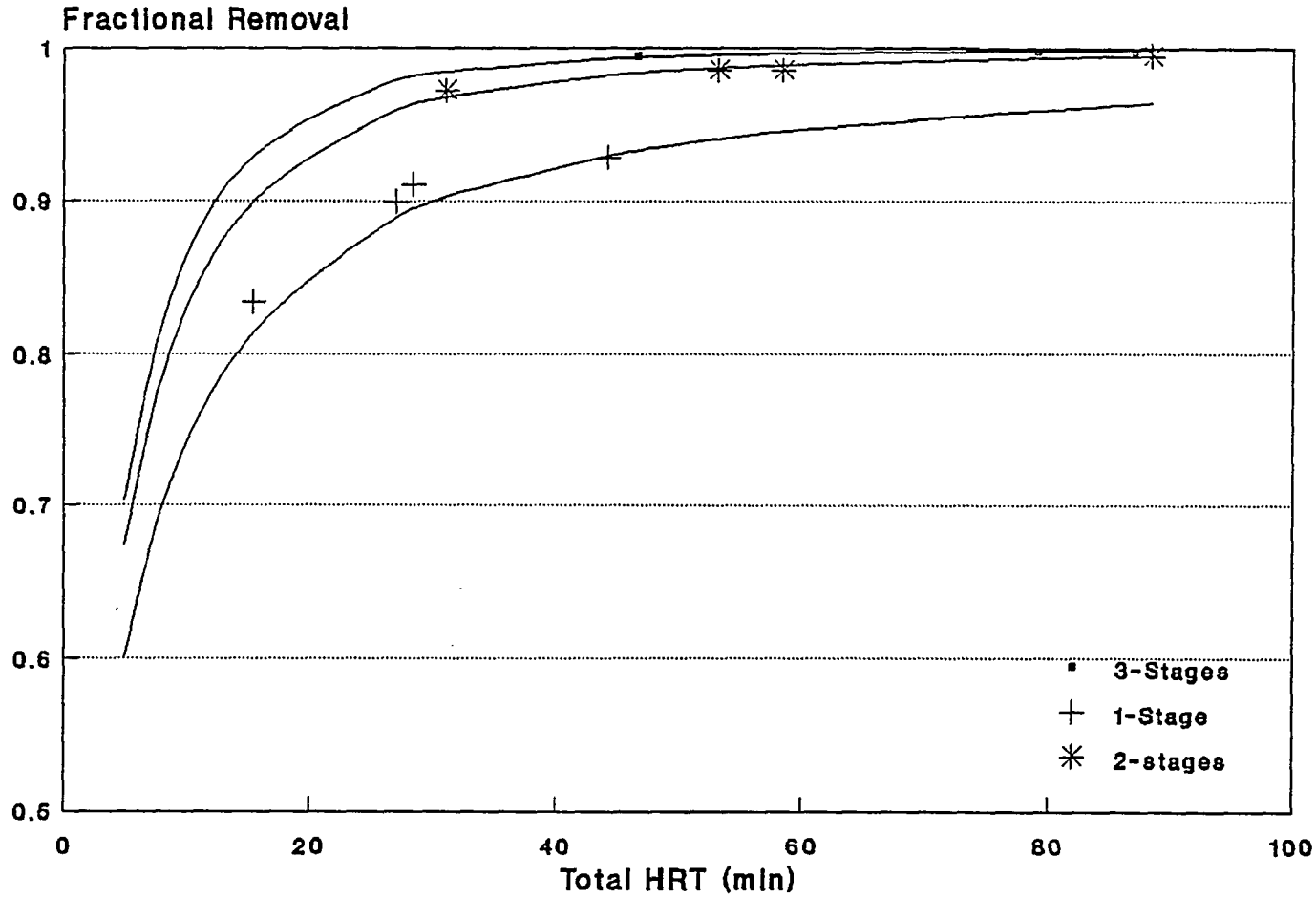


Fig. 5.9 Effects of HRT and Stages on the Removal of TCE (w=12.1 rpm)



•solid lines are theoretical curves calculated with average removal

Fig 5.9b Removal of PCE in Continuous Flow (Rh = 0.68)

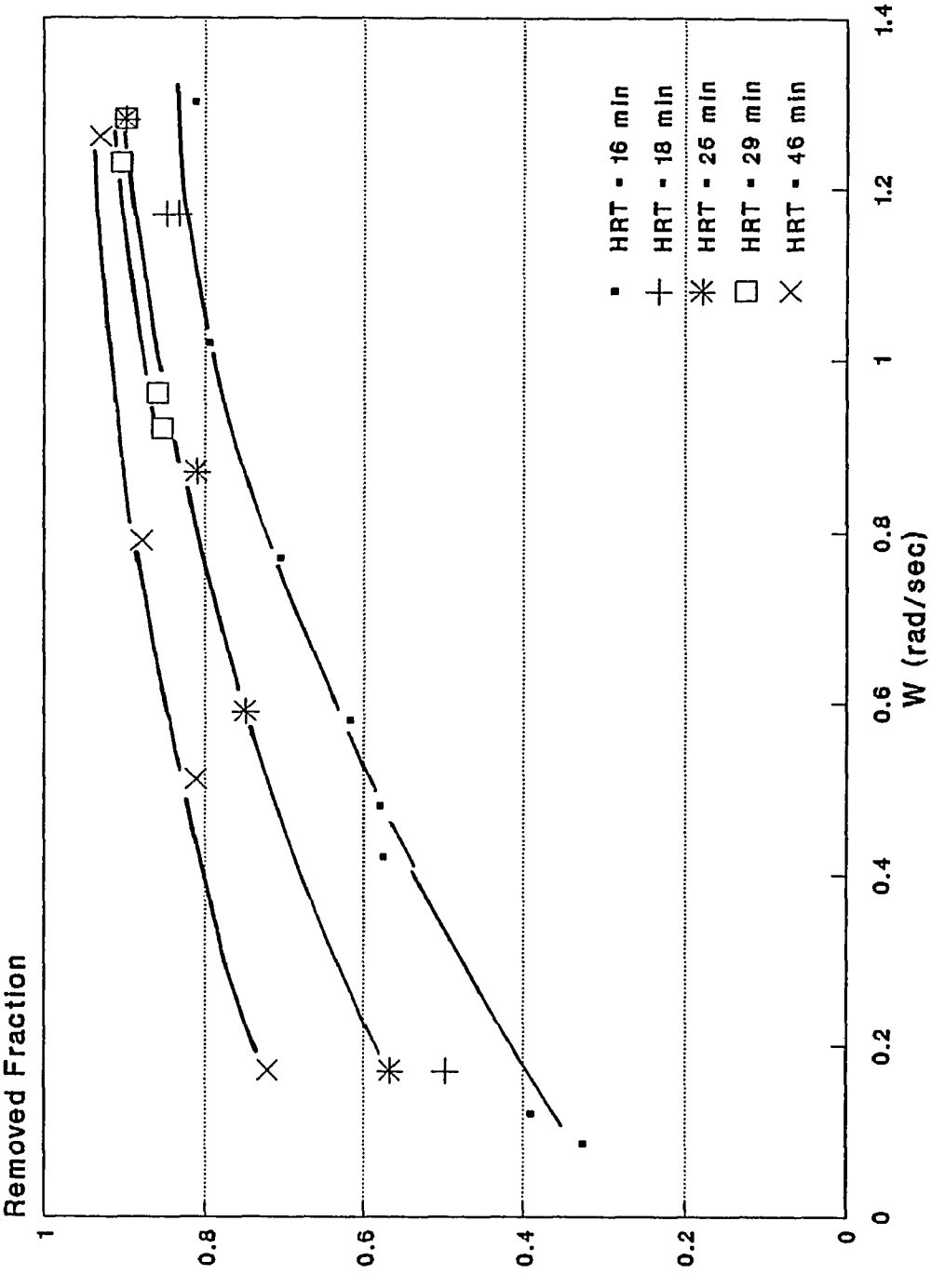


Fig 5.9c Removal of EB in Continuous Flow (Rh = 0.68)

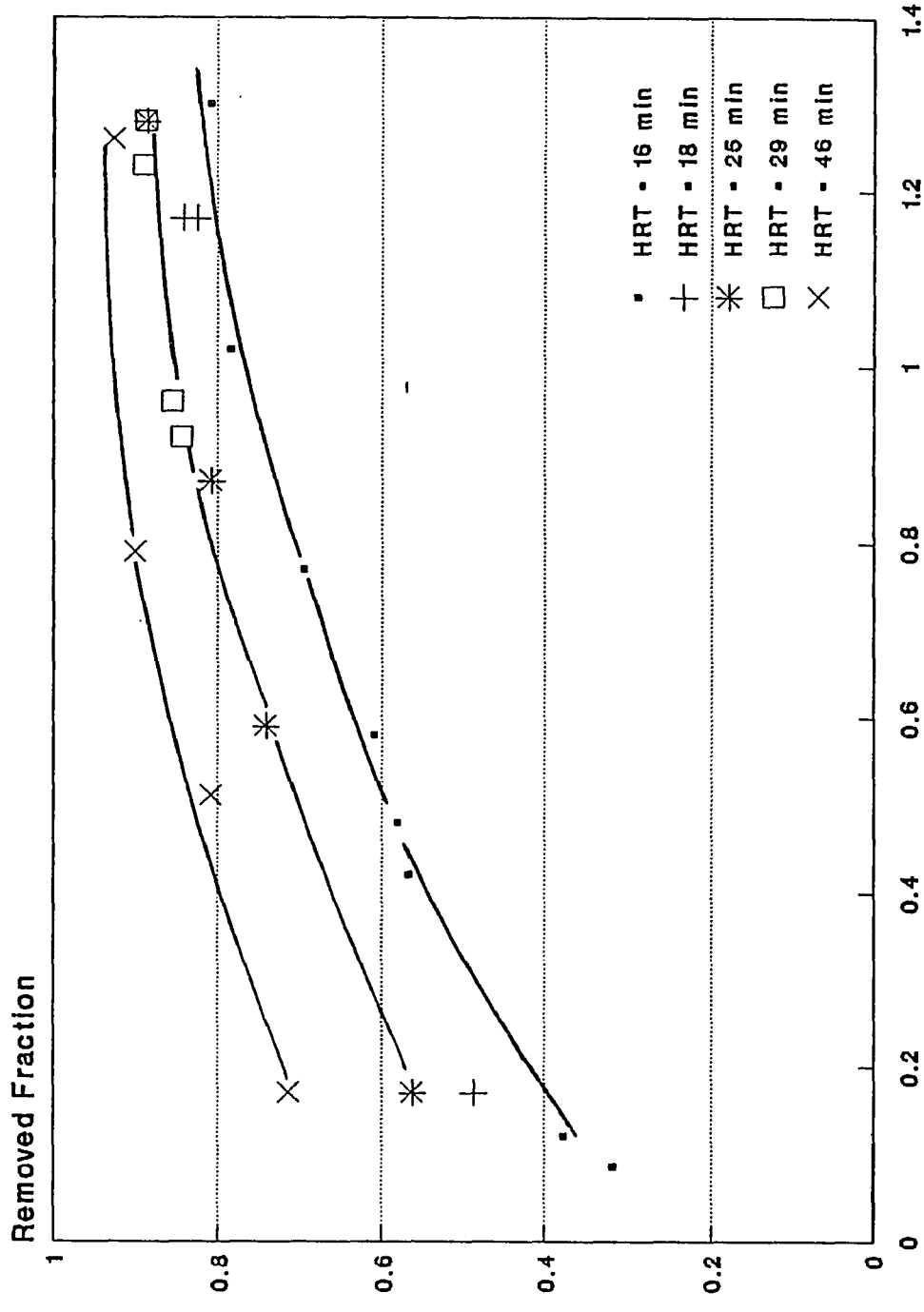


Fig 5.9a Removal of CLF in Continuous Flow (Rh = 0.68)

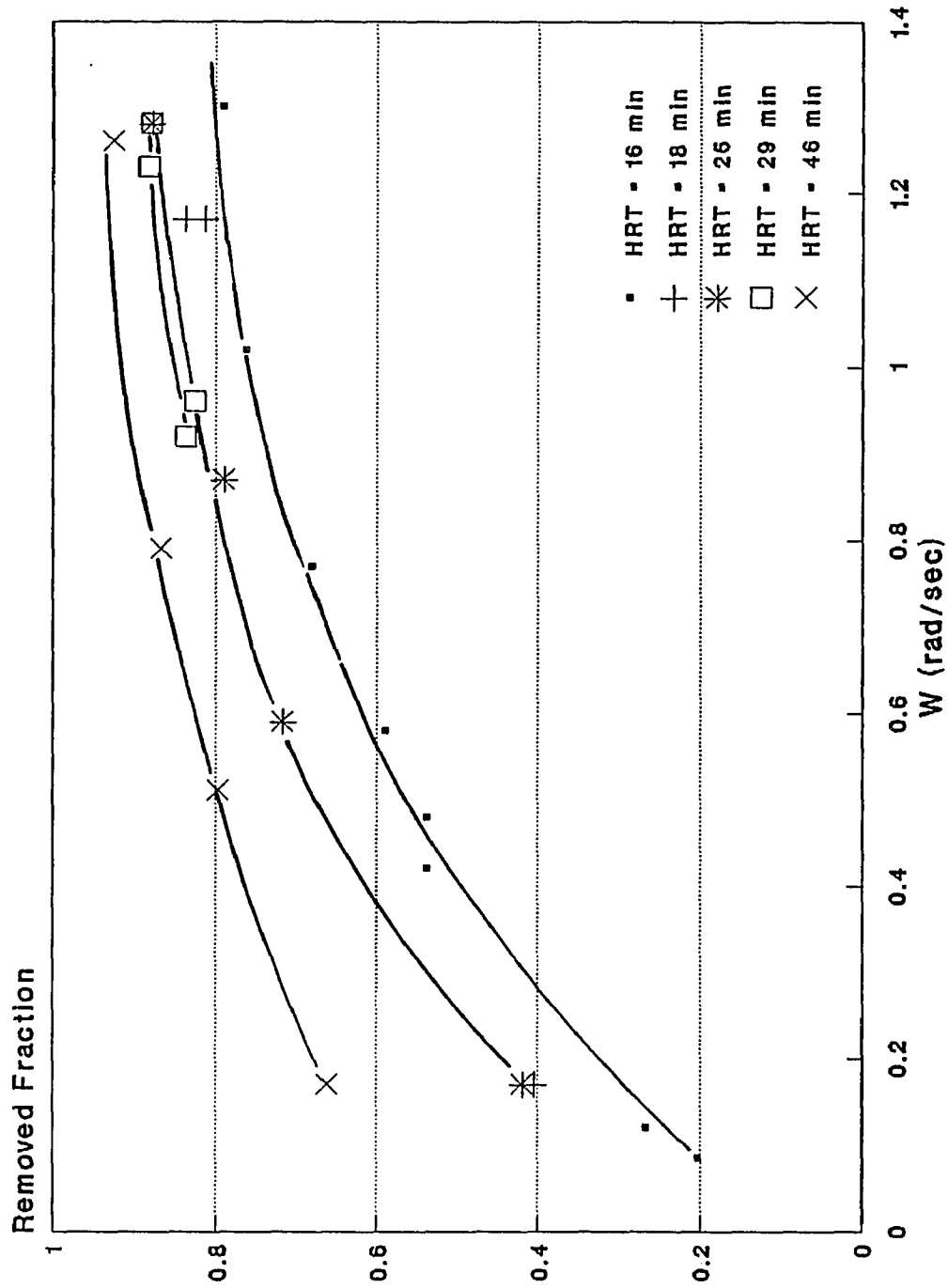
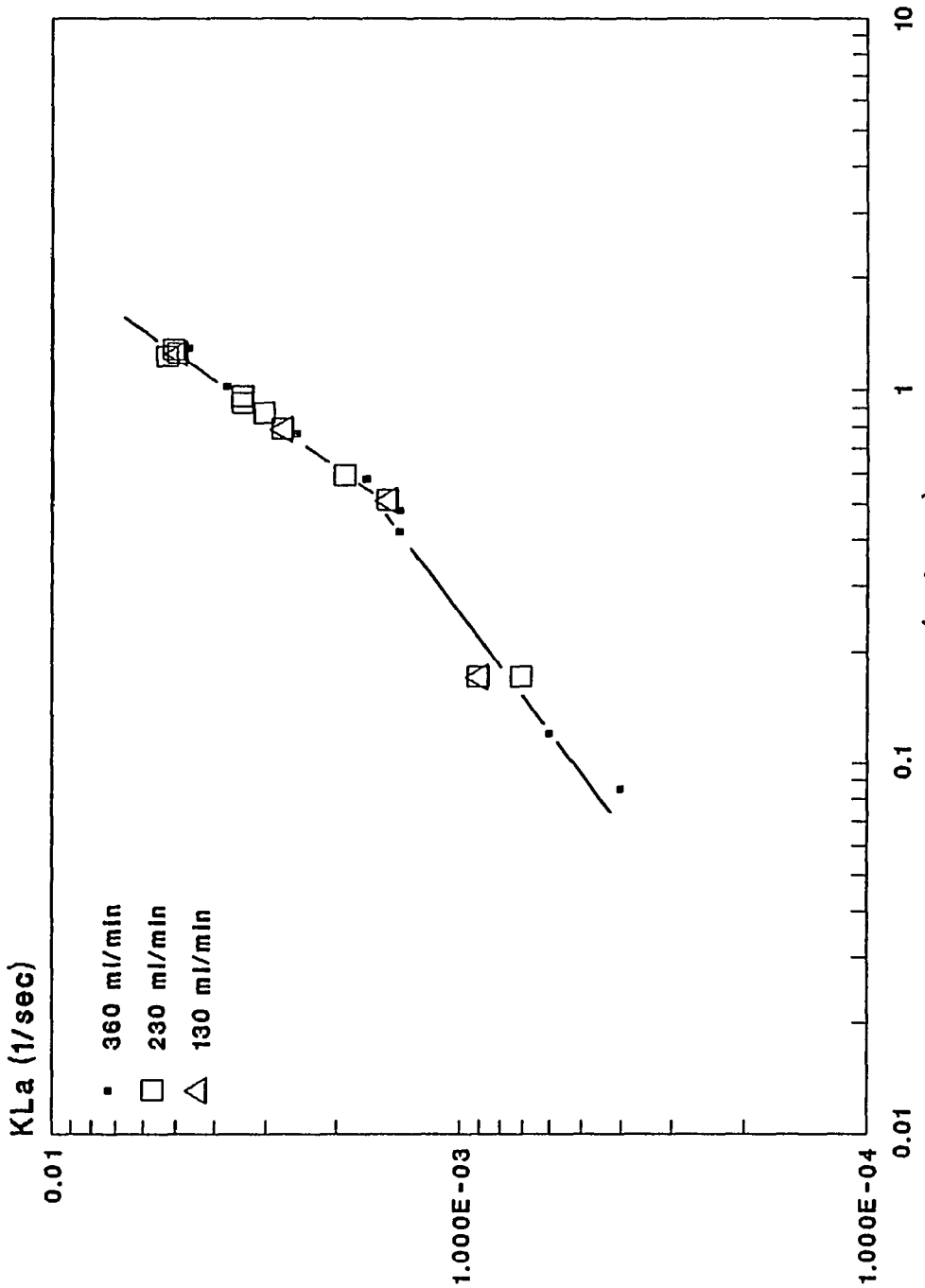


Fig 5.10 KLa of TCE Determined in Continuous Flow Study



5.4 Dependence of K_L on Various Operational Conditions

Two types of troughs are usually used by commercial RDC systems: a steel trough with cylindrical bottom, and a concrete trough with flat bottom. The disks of commercial units are made of polystyrene, polyethylene, polyvinyl chloride or fiberglass. The disk surface may be flat, corrugated, honeycombed or otherwise formed to increase the surface area. Matsuo (1985) reported that different bottom configurations and different disk materials had shown no effect on oxygen transfer. At this time there is no way to quantitatively assess the influence of the disk surface properties on the stripping of VOCs. Since the desorption of VOCs is similar to oxygen absorption, also depending on the hydraulic characteristics of the RDC system, it is assumed that these parameters do not affect the efficiencies of VOC stripping. Kim (1980) proposed that the K_La decreased proportionally with increasing distancebetween disks to a power of 0.86. However, the data presented by Matsuo (1985) showed no significant difference on the observed K_L by varying the spacing between disks. Kim's results are possibly due to the fact that the specific surface area decreases with increasing disk spacing. Limited by the surface configurations of commercial disk, the distance between disks is fixed. From a practical point of view, the impacts of disk spacing, if there are any, have only very minimal importance under normal

operating conditions.

In this study, the influences of different physical operational parameters on VOC stripping and oxygen absorption were studied under a constant water temperature of 20 $\pm$ 3°C. The mass-transfer coefficients, K_L, for different operating conditions were calculated and listed in Table 5.1, 5.2 and 5.4. Data shows that K_L depends significantly on the angular rotational speed and depth of submergence. The data obtained by the batch procedure with bulk water of 4 L (Rh=0.49) were plotted in Fig. 5.11. The data obtained in the continuous flow system (Rh=0.68) were plotted in Fig. 5.12. Fig. 5.13 shows the relationships between oxygen transfer coefficient, rotational speed and depth of disk submergence. When rotational speeds were beyond 3.77 rpm (0.395 rad/sec) and Ri/R less than 0.5, the values of K_L increased proportionally with the rotational speed raised to a power of around 1.42 for all six compounds, as indicated by the slopes of the regression lines in Fig 5.11 and 5.12, with r^2 larger than 0.98. The high correlation coefficient, r^2, confirms the linear relationship between log w and log K_L. The exponents of the rotational speed and r^2 are summarized in Table 5.7 for both batch studies and continuous flow studies. It is noticed from the comparison in Fig 5.11 and 5.12 that the K_L response curves fo VOCs were nearly parallel to that of

oxygen. This observation is consistent with the findings of other studies of VOCs stripping by surface aeration, and indicates the high dependence of K_L on the hydrodynamic properties of the system. The models developed for oxygen transfer should be also applicable to the prediction of the VOCs mass transfer rates. The values of exponents obtained in continuous flow were slightly larger, about 3% to 5%, than those of the batch study. However, the difference was relatively insignificant.

The exponents of the angular velocity dependence for all six compounds were slightly less than 1.5. The less volatile compounds showed a little more dependence on the rotational speeds. For practical purposes the transfer coefficients of the compounds can be considered approximately proportional to the angular velocity to the power of 1.42. According to Eq.3.16 given by Bintanja (1974), and Figure 3.2, when $\delta/\sqrt{Dt_a} > 1.7$, $K_L/2\sqrt{D/\pi t_a}$ is independent to $\delta/\sqrt{Dt_a}$ and has a constant value of 1.0. Eq.3.16 is then reduced to Eq.3.31. Eq. 3.31 can also be derived with the aid of the penetration theory (Bird, 1960). It represents the region of high rotational speeds, i.e. short air-water contact time, at which the penetration model of Higbie holds. It was assumed that the thickness of the liquid film is infinite in comparison with the depth of diffusion of oxygen or VOC near the air-water interface. K_L is a function of diffusivity and contact time and is independent

of the thickness of the liquid film on the disks. Under ideal conditions, K_L should be proportional to rotational speed to a power of 0.5 (Eq.3.31). The oxygen transfer data obtained by Bintanja (1975) from a pilot plant with flat disks of 30 cm in radius showed the mass transfer coefficient to vary proportionally with the square root ofthe angular velocity at higher rotational speeds (w > 30 rpm). Yamane (1972) reported that the Higbie model held for mean contact time between 0.2 and 1.5 seconds (rotational speeds of 250 and 33 rpm) for an RDC system with disks of 4 cm in radius. The maximum rotational speed in this study was about 18 rpm which were far below the critical speeds defined by Yamane and Bintanja. Higher speeds over 18 rpm was not possible because the water film was detached as sprays from the disk surface by centrifugal force, and the bulk water was splashed out from the trough by turbulent agitation.

At low rotational speeds, if the exposure time is long enough, the water film becomes highly saturated. Higbie's assumption of diffusion into a semi-infinite medium is no longer valid because of the finite thickness of the water film on the disk. The rotating disks act as a pump replacing water in the trough with oxygen saturated or VOC free water from the disks. This is the situation when /Dta < 0.8. The Eq. 3.16 is then reduced to Eq. 3.32, the K_L being proportional to rotational speed to a power of

1.5. The values of K_L as a function of w will fit a faint S-shape curve, with a more concave curve at lower rotational speeds, and a more convex curve in the region of high rotational speed. The exponents of the rotational speed obtained in this study inferred that a high degree of saturation might exist in the liquid film. Although no data has been found in the literature concerning the stripping of VOCs, previous unsteady-state aeration studies gave results similar to those obtained from this study. The data presented by Bantanja (1975) showed that the oxygen transfer rate coefficient depended on rotational speeds to a power of 1.3 in the range of 7.5 to 30 rpm, and Ri/R = 0.43. The power exponent of oxygen obtained in this study was 1.36 which was very close to Bantanja's results.

According to Eq. 3.16, k_L is a function of molecular diffusivity, exposure time (contact time), and the thickness of the water film on the exposed disk. The mean exposure time, t_a, is expressed as Eq. 3.26a which was derived geometrically from the ratio of exposed area/total wetted area. Yamane (1972) and Bintanja (1975) proposed much more sophisticated equations as Eq. 3.26b, c, and d on the basis of mean mass transfer rate. All four equations give very similar numerical results, as indicated in Table 3.1. Therefore, it virtually makes no difference which equation has been applied. In order to simplify the

correlation between k_L and operational parameters, Eq. 3.26a was applied throughout the data interpretation.

The thickness of the liquid film on the discs was not directly measured in this study, owing to the relatively complicated surface configurations of the disks. Eq. 3.23 was used to evaluate the thickness of the water film on the exposed disks. This equation was theoretically derived and experimentally confirmed by Zeevalkink (1978) from flat disk RDC systems. The thickness of the liquid film is a function of angular speed, disc radius, depth of immersion and viscosity. This equation was derived by the analogy between disks' rotation and flat plates withdrawn from water with constant velocity (Schlichting, 1960). A similar approach was utilized by Famularo (1978) who applied the equation derived by Levich (1968) of a flat plate withdrawn vertically from a quiescent liquid. Famularo proposed that

$$\delta = 6.85 v^{2/3} + \alpha$$

v is the withdrawal velocity, which is calculated as the vertical component of the tangent velocity at two-thirds of disk radius, and hi is an arbitrary value representing the surface irregularities of the disks. Another approach employed the equations derived with boundary layer theory (Schliching, 1968) and proposed that the film thickness was proportional to rotational speeds to the power of -0.5 and

-1 for turbulent and laminar flow respectively (Ouano, 1981). However, these equations seem to be more appropriate for the water film of fully submerged disks. Zeevalkink's equation has been considered more capable of predicting the water film thickness on the disk.

It is noticed from Eq. 3.26a and 3.27 that both exposure time and water film thickness are influenced by the depth of submergence. The effects of depth of submergence on oxygen transfer were plotted in Fig 5.13. This graph shows the tendency of increasing k_L values with increasing depths. This observation was consistent with the data reported by Zeevalkink (1979) and Matsuo (1985). In the case of saturated water film mass transfer, Eq. 3.32 can be theoretically expressed as Eq. 3.34. The experimentally determined mass transfer coefficient is then correlated with rotational speed, depth of submergence, and disc radius as:

$$K_L = C\, w^{1.5}\, R^{0.5}\, F(\frac{R_i}{R}) \tag{5.1}$$

where C = constant, Ri = distance between the shaft and water surface (m), R = radius of the disc (m), w = rotational speed of the disc (1/sec), F(Ri/R) = function of Ri/R. Function F (Eq. 3.35) is represented graphically in Fig 5.14 with the dimensionless parameter Ri/R which is equal to (1 - Rh). With the assistance of polynomial

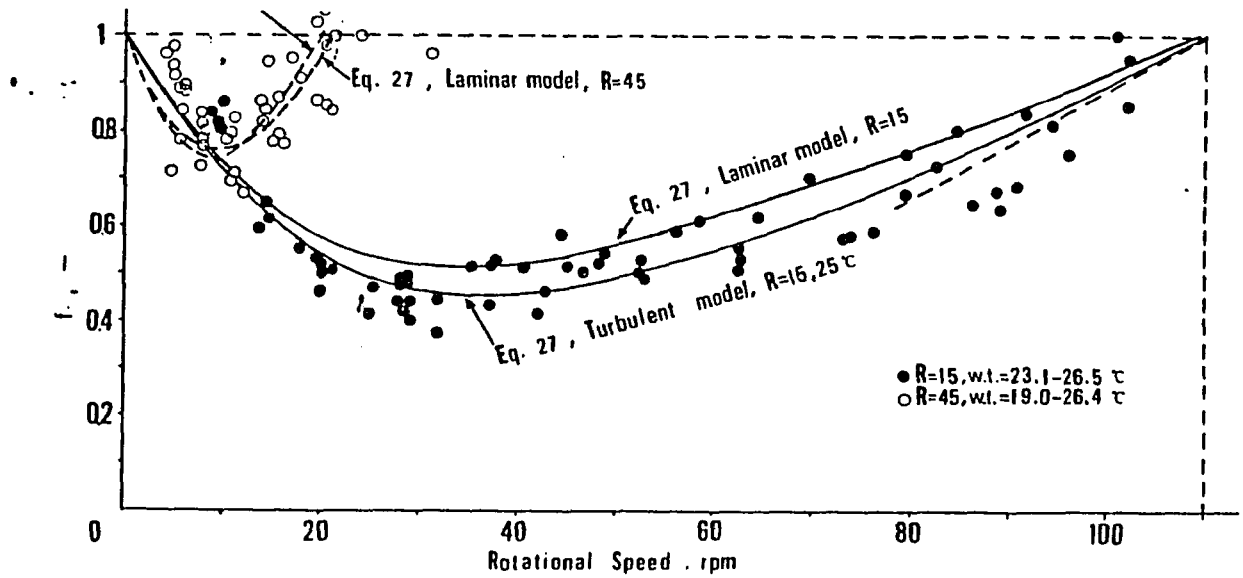
regression, the function is written as Eq. 5.2 to simplify the calculation procedure.

$$F(x) = 2 - 2.63x + 5.249x^2 - 10.07x^3 + 10.68x^4 - 4.788x^5 \qquad (5.2)$$

The constants C of Eq. 5.1 were calculated and plotted in Fig 5.15, 5.16 and 5.17 for batch study, continuous flow study and batch aeration study (various Rh) respectively. As indicated in these graphs, the C-values were reasonably independent of the rotational speeds and depth of submergence (represented by Ri/R or 1-Rh) when rotational speeds were higher than 0.4 rad/sec. At lower rotational speeds, the C-values were roughly inversely proportional to the speeds. The values of C were calculated for the region of higher rotational speeds and listed in Table 5.8 for each selected compounds. The values of C obtained in the continuous flow study were slightly higher than those of the batch study. Such deviations might be due to the extra stripping in the entrance region, deviations of the flow, or due to the difference of surface configurations of the disk at different immersed depths resulting in an inaccurate estimate of the specific contact area. However, all three graphs showed similar trends, and the deviations were limited in a reasonable range.

According to Eq. 3.34, the mass transfer coefficient is dependent on the hydrodynamic properties of the system.

The constant C should be a function of dynamic viscosity of theliquid, density, and the acceleration due to gravity. For water at 20^{o} C, viscosity = $1x10^{-6}$ m^2/s, g = 9.81 m/s^2, and C will be 1.91 x 10^{-5}. When considering the incomplete mixing between water film and bulk water, this value should be even less (Eq. 3.25). Matsuo (1985) proposed that the factor of incomplete mixing, f, could be equated as Eq. 3.28. The theoretically and experimentally determined values of f were graphically presented by Matsuo as functions of rotational speeds and disk radius (see the graph below which is adopted from the same reference)



As indicated in this graph, for an RDC system with 15 cm in disk radius, the values of f decreased with increasing rotational speed to the minimum at about 30 rpm and then increased with increasing speeds to the maximum of 1 at about 110 rpm. With larger disk radius, the values of f approached 1 at lower rotational speeds. Obviously, the

value of C determined in this study did not exhibit similar characteristics.

Matsuo (1985) studied the mixing of water film on flat disks with bulk water by tracer. He pointed out that the water film submerged smoothly into the bulk water and went on without being disturbed during the submergence. The mass transfer between water film and bulk water was controlled by the molecular diffusion process. At the points of emergence, the formation of a strong eddy current contributed to the major mechanism of mixing between the water film and bulk water. In this study, milk and 600 mg/l methylene blue were applied to the disk surface as tracers for visualizing the mixing pattern. Contrary to the observations of Matsuo, it was found that, even at the rotational speed as low as 1 rpm, the tracers mixed with the water quickly after submerged into the bulk water. Mixing by turbulence seemed to control the dispersion of the tracers from the disk surface throughout the submerged cycle. The difference between observations may be explained by the different surface configurations of the disks. Flat disks were used in Matsuo's experiment, while sockets were distributed on the disks' surface in this study. Those sockets increased the surface irregularities which acted as turbulence promoters, enhancing the micro-mixing near the disk surface and the macroscopic turbulence

in the trough. Considering all these observations, it is concluded that the incomplete mixing did not play an important role in this study, and thus f = 1.

For the experimental data of flat disk reported by Bintanja (1975) and Zeevalkink (1979), the corresponding C-values were calculated as 1.97 and 2.1 x 10^{-5} respectively. The C-value of oxygen in this study was found to be 3.107 x 10^{-5}, which was 1.63 times the theoretical value of 1.91 x 10^{-5}, or about 1.5 times that of previous flat disk experiments. It inferred that the configuration of the disk surface used in this study enhanced the interphase mass transfer. The basic model was based on several assumptions as described in Chapter 3. Apparently some of them are not fully justified. One significant deviation is associated with the assumption of an uniform stagnant film existing on the exposed disk surface. The film thickness is a function of the velocity profile on the disk. The true thickness varies from point to point over the disk surface. The uniform thickness is actually not possible. The variation of the directions of the net forces causes the water film to be unstable. Visual observation showed that the water film at higher rotational speeds was not at rest with respect to the disk but flowed under the influence of gravity and rotational acceleration. The sockets on the disk and roughness of the surface tends to bring up more water at the points of emergence than a flat disk. The

excessive water will then flow downward during the exposure cycle. On one hand, the roughness and geometrical irregularities tend to increase the water film thickness, and, on the other hand, the unstable water film can grow and raise the turbulence level in the film. The presence of turbulence and surface convection in the water film may increase the transfer coefficient considerably (Grossman, 1986). Compared with the data obtained from flat disks the enhancement will be 1.5 to 1.6 for oxygen transfer. Since there is no data available in the literature concerning the stripping of VOCs in RDC systems, no comparison was made for VOCs.

As shown in Fig 5.11 and 5.12, in the region of low rotational speeds, the dependence of K_L on the rotational speed is relatively weak. The values of K_L were higher than the values predicted by Eq 3.32 or 3.34 which represented a water film controlled mass transfer mechanism. The difference between observed values and predicted values became more obvious with decreasing rotational speeds. This phenomenon resulted in the increase of the C-value with decreasing rotational speeds. Such observations inferred that a mechanism other than water film transfer might dominate at lower rotational speeds.

Eq. 3.9 indicates that the overall mass transfer is

the sum of the water film transfer and the direct transfer to the bulk water surface. The mass transfer between air and bulk water surface has been neglected by most authors, due to its small surface area compared with the vast disk area. However, at low rotational speeds, the turn over rate of the water film on the disks becomes very small. The relative importance of the water surface transfer becomes more pronounced. The mass transfer rate constant, denoted as $K_L a_l$, is dependent upon the turbulence at the air/water interface. Rittmann (1983) used a propeller mixer to simulate the oxygen absorption on the turbulent bulk water surface and found that the values of $K_L a_l$ were proportional to rotational speed to the power of 0.63. Naturally, the propeller mixer cannot exactly simulate the true water surface transfer in an RDC system. To evaluate the relative importance of water surface transfer, Eq. 3.9 was rearranged as

$$K_L = K_{Ll}\frac{a_l}{a} + K_{Ld}\frac{a_d}{a} \tag{5.3}$$

where a_l/a and a_d/a are the fractions of total contact area contributed by water surface and exposed disk area respectively, and whose values can be found in Table 5.5. K_{Ld} is equated as Eq. 5.1. Since the values of C showed relatively weak dependence on the depth of submergence (see Fig 5.17), the K_{Ll} was considered to be a function of rotational speed only. Eq.5.3 was then equated as

$$K_L = C_l w^{n_1} + C_d R^{0.5} F(R_i/R) w^{n_d} \tag{5.4}$$

Since four unknowns were involved, iterative trial-error searches were employed for the minimal MSD. the data from the batch study were correlated with the operational parameters as the following:

	$K_{L1}(a_1/a)$	$K_{Ld}(a_d/a)$
O_2	$(59)(0.05)w^{0.33}$	$(27.0)(0.95)f(\frac{R_i}{R})R^{0.5}w^{1.56}$
TCE	$(47)(0.05)w^{0.33}$	$(16.0)(0.95)f(\frac{R_i}{R})R^{0.5}w^{1.53}$
PCE	$(48)(0.05)w^{0.33}$	$(15.3)(0.95)f(\frac{R_i}{R})R^{0.5}w^{1.53}$
EB	$(48)(0.05)w^{0.33}$	$(15.3)(0.95)f(\frac{R_i}{R})R^{0.5}w^{1.53}$
CLF	$(39)(0.05)w^{0.33}$	$(15.3)(0.95)f(\frac{R_i}{R})R^{0.5}w^{1.53}$
TCEN	$(49)(0.05)w^{0.33}$	$(15.3)(0.95)f(\frac{R_i}{R})R^{0.5}w^{1.53}$

(5.5)

of K_L were listed in Table 5.10 and 5.11. The average errors of the prediction were within 5% for most of the data. As indicated in these two tables, the K_{L1} predominated in the region of low rotational speeds. As rotational speed increased, $K_L a_1$ became a smaller fraction of K_L. Thus, with the increasing speeds, direct transfer to the bulk water surface becomes less significant, which is consistent with previous observations reported by

Rittmann (1983). Based on these estimates, the K_L of oxygen and TCE were plotted in Fig 5.18 and 5.19 as examples to illustrate the extent of each mechanism. For rotational speeds less than 0.5 rpm direct transfer to the bulk water surface dominated. The water film mass transfer contributed less than 10% of the total mass transfer in this region. For rotational speeds in the range of 0.5 to 2.5 rpm, direct transfer still dominated, but the film transfer played a more important role. When rotational speeds were higher than 4 rpm, K_{Ld} controlled the overall mass transfer process. Rittmann (1985) estimated that the water surface mass transfer rate was approximately proportional to the rotational speed (in terms of power input) to a power of 0.63. The dependence of the water surface transfer on rotational speed was relatively weak as estimated in this study. This may be explained by the fact that the irregularities of the disk surface promoted the turbulent level and thus decreased the dependence on power input. Although this estimation method provided a crude approximation, the results revealed that the significance of water surface may not be neglected under some operational conditions.

Table 5.7 The Exponent of Rotational Speed and r^2

Compounds	Exponent n			
	Batch	r^2	Cont. Flow	r^2
Oxygen	1.363	0.994	1.41	0.993
Trichloroethylene	1.390	0.980	1.43	0.986
Tetrachloroethylene	1.405	0.982	1.45	0.986
Ethylbenzene	1.405	0.988	1.40	0.987
1,1,1-Trichloroethane	1.439	0.989	1.49	0.990
Chloroform	1.460	0.984	1.38	0.981

Table 5.8 Calculated Values of Constant C

Compounds	Constant C (10^{-5})	
	Batch	Cont. Flow
Oxygen	3.107	3.64
Trichloroethylene	1.839	2.06
Tetrachloroethylene	1.757	2.08
Ethylbenzene	1.758	2.04
1,1,1-Trichloroethane	1.743	2.01
Chloroform	1.661	1.87

Table 5.10 Data for Oxygen Transfer (Batch Study)

Rotational Speed, w (rpm)	(rad/sec)	Specific Area, a (1/m)	Rh i.e. (H/R)	Ri/R i.e. (1-Rh)	$F(\frac{R_i}{R})$	Water Temp. (°C)	obs.KLa 10-3 (1/sec)	obs.KL 10-6 (m/sec)	(al/a)KLl 10-6 (m/sec)	%KL	(ad/a)KLd 10-6 (m/sec)	predicted. KL 10-6 (m/sec)	% error
0.23	0.024	290	0.49	0.51	1.248	30	0.32	1.11	0.980	95.9	0.041	1.021	-0.083
0.36	0.038	290	0.49	0.51	1.248	25	0.33	1.14	1.125	93.0	0.085	1.210	0.063
0.78	0.082	290	0.49	0.51	1.248	30	0.50	1.72	1.414	83.5	0.279	1.693	-0.018
0.81	0.085	290	0.49	0.51	1.248	20	0.53	1.83	1.432	82.8	0.298	1.730	-0.053
1.22	0.128	290	0.49	0.51	1.248	30	0.60	2.07	1.619	74.1	0.565	2.184	0.056
2.48	0.260	290	0.49	0.51	1.248	20	0.75	2.59	2.003	54.0	1.706	3.709	0.434
3.48	0.364	290	0.49	0.51	1.248	23	1.45	5.00	2.215	43.4	2.884	5.099	0.020
4.13	0.433	290	0.49	0.51	1.248	20	1.72	5.93	2.334	38.2	3.781	6.115	0.031
5.27	0.552	290	0.49	0.51	1.248	20	2.32	8.00	2.510	31.3	5.522	8.032	0.004
5.70	0.597	290	0.49	0.51	1.248	21	2.53	8.72	2.570	29.2	6.240	8.810	0.010
5.90	0.618	290	0.49	0.51	1.248	20	2.46	8.48	2.597	28.3	6.586	9.183	0.083
6.58	0.689	290	0.49	0.51	1.248	20	3.05	10.52	2.683	25.6	7.804	10.486	-0.003
7.68	0.804	290	0.49	0.51	1.248	25	3.75	12.93	2.810	22.1	9.928	12.738	-0.015
7.90	0.827	290	0.49	0.51	1.248	20	3.87	13.33	2.834	21.5	10.375	13.209	-0.009
9.10	0.953	290	0.49	0.51	1.248	20	4.60	15.86	2.957	18.6	12.944	15.901	0.002
9.93	1.040	290	0.49	0.51	1.248	20	5.62	19.38	3.036	17.0	14.834	17.869	-0.078
11.17	1.170	290	0.49	0.51	1.248	22	5.91	20.38	3.145	15.0	17.826	20.970	0.029
12.03	1.260	290	0.49	0.51	1.248	21	6.57	22.66	3.215	13.8	20.010	23.226	0.025
13.94	1.460	290	0.49	0.51	1.248	20	7.60	26.21	3.361	11.8	25.181	28.541	0.089
15.25	1.597	290	0.49	0.51	1.248	30	9.70	33.45	3.452	10.7	28.962	32.415	-0.031
15.64	1.638	290	0.49	0.51	1.248	25	10.38	35.78	3.479	10.4	30.131	33.609	-0.061
17.19	1.800	290	0.49	0.51	1.248	20	11.28	38.90	3.579	9.3	34.906	38.485	-0.011
0.27	0.028	343	0.39	0.61	1.139	25	0.33	0.96	1.026	95.5	0.048	1.074	0.117
0.45	0.047	343	0.39	0.61	1.139	20	0.47	1.37	1.199	91.7	0.108	1.307	-0.046
0.82	0.085	343	0.39	0.61	1.139	20	0.59	1.72	1.434	83.9	0.274	1.708	-0.007
1.64	0.172	343	0.39	0.61	1.139	25	0.93	2.71	1.769	68.4	0.817	2.587	-0.046
2.44	0.256	343	0.39	0.61	1.139	20	1.18	3.44	1.993	56.7	1.520	3.513	0.021
4.89	0.512	343	0.39	0.61	1.139	20	2.39	6.95	2.454	35.4	4.482	6.936	-0.003
6.77	0.709	343	0.39	0.61	1.139	20	3.69	10.76	2.706	26.7	7.442	10.148	-0.057
11.75	1.230	343	0.39	0.61	1.139	20	6.62	19.30	3.192	15.4	17.589	20.781	0.077
12.60	1.319	343	0.39	0.61	1.139	22	7.50	21.85	3.260	14.3	19.614	22.874	0.047
14.10	1.477	343	0.39	0.61	1.139	20	9.30	27.11	3.372	12.6	23.400	26.773	-0.013
14.40	1.508	343	0.39	0.61	1.139	22	9.16	26.71	3.393	12.3	24.171	27.564	0.032
15.95	1.670	343	0.39	0.61	1.139	20	10.00	29.15	3.499	11.0	28.342	31.841	0.092
16.50	1.728	343	0.39	0.61	1.139	19	11.30	32.94	3.535	10.6	29.892	33.427	0.015

Table 5.10 Data for Oxygen Transfer (Batch Study)

Rotational Speed, w (rpm)	(rad/sec)	Specific Area, a (1/m)	Rh i.e. (H/R)	Ri/R i.e. (1-Rh)	F( )	Water Temp. (C)	obs.KLa 10-3 (1/sec)	obs.KL 10-6 (m/sec)	(al/a)KLl 10-6 (m/sec)	%KL	(ad/a)KLd 10-6 (m/sec)	predicted. KL 10-6 (m/sec)	% error
0.37	0.039	415	0.28	0.72	1.012	25	0.52	1.26	1.134	94.0	0.072	1.205	-0.042
0.45	0.047	415	0.28	0.72	1.012	20	0.60	1.45	1.199	92.6	0.096	1.295	-0.104
1.22	0.128	415	0.28	0.72	1.012	25	0.79	1.90	1.619	78.0	0.458	2.077	0.091
1.63	0.171	415	0.28	0.72	1.012	25	1.07	2.58	1.766	71.1	0.720	2.486	-0.036
3.06	0.320	415	0.28	0.72	1.012	20	1.65	3.98	2.131	52.7	1.912	4.044	0.017
4.87	0.510	415	0.28	0.72	1.012	20	2.76	6.65	2.451	36.3	3.957	6.408	-0.036
5.85	0.613	415	0.28	0.72	1.012	20	2.98	7.17	2.590	32.9	5.272	7.863	0.096
6.49	0.680	415	0.28	0.72	1.012	25	3.59	8.65	2.672	30.1	6.198	8.871	0.025
7.73	0.810	415	0.28	0.72	1.012	20	4.42	10.65	2.816	25.7	8.143	10.959	0.029
8.13	0.851	415	0.28	0.72	1.012	20	5.00	12.05	2.858	24.5	8.795	11.654	-0.033
9.31	0.975	415	0.28	0.72	1.012	20	5.62	13.54	2.977	21.5	10.874	13.852	0.023
10.46	1.095	415	0.28	0.72	1.012	20	6.56	15.81	3.083	19.1	13.033	16.116	0.020
11.34	1.188	415	0.28	0.72	1.012	20	6.56	15.81	3.159	17.6	14.800	17.959	0.136
12.61	1.320	415	0.28	0.72	1.012	20	7.90	19.04	3.261	15.7	17.444	20.705	0.088
13.79	1.444	415	0.28	0.72	1.012	20	9.00	21.69	3.350	14.3	20.067	23.417	0.080
14.42	1.510	415	0.28	0.72	1.012	22	10.10	24.34	3.395	13.6	21.516	24.911	0.024
14.60	1.529	415	0.28	0.72	1.012	20	10.60	25.55	3.408	13.4	21.940	25.347	-0.008
15.47	1.620	415	0.28	0.72	1.012	25	10.52	25.35	3.467	12.6	24.011	27.478	0.084
15.80	1.655	415	0.28	0.72	1.012	20	11.70	28.19	3.489	12.3	24.825	28.314	0.004
16.52	1.730	415	0.28	0.72	1.012	20	12.20	29.40	3.536	11.7	26.602	30.138	0.025
0.25	0.026	530	0.15	0.85	0.827	20	0.77	1.45	1.405	98.0	0.028	1.433	-0.013
0.41	0.043	530	0.15	0.85	0.827	25	0.83	1.57	1.634	96.4	0.062	1.696	0.083
0.84	0.088	530	0.15	0.85	0.827	20	0.91	1.71	2.026	91.5	0.189	2.214	0.295
1.63	0.171	530	0.15	0.85	0.827	25	1.42	2.68	2.473	82.3	0.532	3.004	0.121
2.26	0.237	530	0.15	0.85	0.827	20	1.73	3.08	2.727	75.5	0.885	3.612	0.173
4.89	0.512	530	0.15	0.85	0.827	20	3.12	5.89	3.436	53.9	2.942	6.377	0.083
6.24	0.653	530	0.15	0.85	0.827	20	4.06	7.66	3.696	46.2	4.301	7.997	0.044
8.77	0.918	530	0.15	0.85	0.827	20	5.66	10.68	4.094	35.9	7.317	11.410	0.068
11.59	1.214	530	0.15	0.85	0.827	20	7.90	14.91	4.452	28.2	11.315	15.767	0.058
14.61	1.530	530	0.15	0.85	0.827	20	11.18	21.09	4.772	22.7	16.233	21.005	-0.004
16.08	1.684	530	0.15	0.85	0.827	20	12.91	24.36	4.911	20.7	18.853	23.764	-0.024

Table 5.11 Prediction of Mass Transfer Coefficient (Batch Study Data For VOCs)

Rotational Speed, w (rpm)	(rad/s)	Rh (H/R)	KLa 10-3 1/s	KL 10 -6 m/s obs.	pred.	error	Trichloroethylene KLl 10-3 1/s	% KL	KLd 10-3 1/s	% KL
1.62	0.170	0.49	0.63	2.17	1.93	-0.111	1.38	71.5	0.55	28.5
3.50	0.367	0.49	0.87	2.99	3.52	0.178	1.74	49.4	1.78	50.6
4.49	0.470	0.49	1.26	4.34	4.48	0.031	1.87	41.8	2.60	58.2
7.76	0.813	0.49	2.33	8.03	8.23	0.025	2.21	26.8	6.02	73.2
10.80	1.131	0.49	3.30	11.38	12.42	0.092	2.44	19.6	9.98	80.4
11.22	1.175	0.49	3.99	13.76	13.05	-0.052	2.47	18.9	10.58	81.1
11.83	1.239	0.49	4.30	14.83	13.98	-0.057	2.51	17.9	11.48	82.1
11.94	1.250	0.49	4.03	13.90	14.15	0.018	2.51	17.8	11.63	82.2
12.35	1.293	0.49	4.45	15.34	14.79	-0.036	2.54	17.2	12.25	82.8
12.80	1.340	0.49	4.60	15.86	15.50	-0.023	2.57	16.5	12.94	83.5
13.75	1.440	0.49	4.92	16.97	17.07	0.006	2.62	15.4	14.45	84.6
15.01	1.572	0.49	5.56	19.17	19.21	0.002	2.69	14.0	16.52	86.0
15.24	1.596	0.49	5.80	20.00	19.61	-0.019	2.70	13.8	16.91	86.2
16.03	1.679	0.49	6.50	22.41	21.02	-0.062	2.75	13.1	18.27	86.9
16.42	1.720	0.49	6.50	22.41	21.72	-0.031	2.77	12.7	18.96	87.3
16.49	1.727	0.49	6.10	21.03	21.84	0.039	2.77	12.7	19.08	87.3
17.57	1.840	0.49	6.40	22.07	23.84	0.080	2.82	11.8	21.02	88.2
17.86	1.870	0.49	6.60	22.76	24.38	0.071	2.84	11.6	21.55	88.4
17.99	1.884	0.49	7.10	24.48	24.63	0.006	2.84	11.5	21.79	88.5
13.82	1.447	0.39	5.23	15.25	15.91	0.043	2.63	16.5	13.28	83.5
13.37	1.400	0.34	5.37	14.21	14.92	0.050	2.60	17.4	12.32	82.6
11.55	1.209	0.28	4.90	11.81	11.45	-0.030	2.49	21.7	8.96	78.3
12.49	1.308	0.28	6.10	14.70	12.66	-0.139	2.55	20.1	10.11	79.9
14.11	1.478	0.28	6.02	14.51	15.36	0.059	3.17	20.6	12.19	79.4
12.61	1.320	0.15	6.50	12.26	11.78	-0.040	3.58	30.4	8.20	69.6
5.83	0.611	0.15								

Table 5.11 Prediction of Mass Transfer Coefficient (Batch Study Data For VOCs)

Rotational Speed, w (rpm)	(rad/s)	Rh (H/R)	Tetrachloroethylene KLa 10-3 1/s	KL 10 -6 m/s obs.	pred.	error	KLl 10-3 1/s	% KL	KLd 10-3 1/s	% KL
1.62	0.170	0.49	0.64	2.21	1.89	-0.143	1.40	73.8	0.50	26.2
3.50	0.367	0.49								
4.49	0.470	0.49	1.23	4.24	4.32	0.019	1.89	43.8	2.43	56.2
7.76	0.813	0.49	2.25	7.76	7.94	0.023	2.23	28.1	5.71	71.9
10.80	1.131	0.49	3.35	11.55	12.01	0.040	2.46	20.5	9.55	79.5
11.22	1.175	0.49								
11.83	1.239	0.49	4.08	14.07	13.54	-0.037	2.53	18.7	11.01	81.3
11.94	1.250	0.49	3.98	13.72	13.70	-0.002	2.54	18.5	11.16	81.5
12.35	1.293	0.49								
12.80	1.340	0.49	4.60	15.86	15.03	-0.052	2.59	17.2	12.44	82.8
13.75	1.440	0.49								
15.01	1.572	0.49	5.58	19.24	18.68	-0.029	2.72	14.6	15.96	85.4
15.24	1.596	0.49	5.90	20.34	19.07	-0.062	2.73	14.3	16.34	85.7
16.03	1.679	0.49	6.40	22.07	20.46	-0.073	2.77	13.6	17.69	86.4
16.42	1.720	0.49	6.54	22.55	21.16	-0.062	2.79	13.2	18.37	86.8
16.49	1.727	0.49								
17.57	1.840	0.49								
17.86	1.870	0.49	6.70	23.10	23.79	0.030	2.87	12.0	20.92	88.0
17.99	1.884	0.49								
13.82	1.447	0.39	5.90	17.20	15.45	-0.102	2.65	17.2	12.80	82.8
13.37	1.400	0.34	5.50	14.55	14.49	-0.004	2.63	18.1	11.86	81.9
11.55	1.209	0.28	4.93	11.88	11.11	-0.065	2.51	22.6	8.59	77.4
12.49	1.308	0.28	6.50	15.66	12.29	-0.215	2.57	20.9	9.72	79.1
14.11	1.478	0.28	6.06	14.60	14.43	-0.012	2.67	18.5	11.76	81.5
12.61	1.320	0.15	6.10	11.51	11.50	-0.001	3.61	31.4	7.88	68.6
5.83	0.611	0.15	3.00	5.66	5.24	-0.074	2.87	54.7	2.37	45.3

Table 5.11 Prediction of Mass Transfer Coefficient (Batch Study Data For VOCs)

Rotational Speed, w (rpm)	(rad/s)	Rh (H/R)	KLa 10-3 1/s	KL 10-6 m/s obs.	pred.	error	Ethylbenzene KLl 10-3 1/s	% KL	KLd 10-3 1/s	% KL
1.62	0.170	0.49	0.64	2.21	1.92	-0.130	1.40	72.7	0.52	27.3
3.50	0.367	0.49								
4.49	0.470	0.49	1.03	3.55	4.38	0.232	1.89	43.3	2.48	56.7
7.76	0.813	0.49	2.10	7.24	7.97	0.101	2.23	28.0	5.74	72.0
10.80	1.131	0.49	2.92	10.07	11.98	0.190	2.46	20.6	9.51	79.4
11.22	1.175	0.49								
11.83	1.239	0.49	4.03	13.90	13.47	-0.031	2.53	18.8	10.94	81.2
11.94	1.250	0.49	3.80	13.10	13.63	0.040	2.54	18.6	11.09	81.4
12.35	1.293	0.49								
12.80	1.340	0.49	4.30	14.83	14.93	0.007	2.59	17.4	12.33	82.6
13.75	1.440	0.49								
15.01	1.572	0.49	5.45	18.79	18.47	-0.017	2.72	14.7	15.75	85.3
15.24	1.596	0.49	5.70	19.66	18.85	-0.041	2.73	14.5	16.11	85.5
16.03	1.679	0.49	6.10	21.03	20.19	-0.040	2.77	13.7	17.41	86.3
16.42	1.720	0.49	6.40	22.07	20.86	-0.055	2.79	13.4	18.07	86.6
16.49	1.727	0.49								
17.57	1.840	0.49								
17.86	1.870	0.49	6.60	22.76	23.40	0.028	2.87	12.2	20.54	87.8
17.99	1.884	0.49								
13.82	1.447	0.39	5.50	16.03	15.31	-0.045	2.65	17.3	12.66	82.7
13.37	1.400	0.34	5.60	14.81	14.37	-0.030	2.63	18.3	11.74	81.7
11.55	1.209	0.28	4.85	11.69	11.06	-0.054	2.51	22.7	8.54	77.3
12.49	1.308	0.28	6.03	14.53	12.21	-0.160	2.57	21.1	9.64	78.9
14.11	1.478	0.28	6.02	14.51	14.04	-0.032	2.67	19.0	11.37	81.0
12.61	1.320	0.15	6.40	12.08	11.43	-0.053	3.61	31.6	7.82	68.4
5.83	0.611	0.15	2.90	5.47	5.27	-0.036	2.87	54.4	2.41	45.6

Table 5.11 Prediction of Mass Transfer Coefficient (Batch Study Data For VOCs)

Rotational Speed, w (rpm)	(rad/s)	Rh (H/R)	KLa 10-3 m/s	1,1,1 Trichloroethane KL 10 -6 m/s obs.	pred.	error	KLl 10-3 1/s	% KL	KLd 10-3 1/s	% KL
1.62	0.170	0.49								
3.50	0.367	0.49								
4.49	0.470	0.49	0.86	2.97	4.02	0.355	1.53	38.2	2.48	61.8
7.76	0.813	0.49	1.90	6.55	7.55	0.152	1.81	24.0	5.74	76.0
10.80	1.131	0.49	3.00	10.34	11.51	0.113	2.00	17.4	9.51	82.6
11.22	1.175	0.49								
11.83	1.239	0.49	4.15	14.31	12.99	-0.092	2.05	15.8	10.94	84.2
11.94	1.250	0.49	3.90	13.45	13.15	-0.022	2.06	15.7	11.09	84.3
12.35	1.293	0.49								
12.80	1.340	0.49	4.30	14.83	14.43	-0.027	2.10	14.6	12.33	85.4
13.75	1.440	0.49								
15.01	1.572	0.49								
15.24	1.596	0.49	5.77	19.90	18.33	-0.079	2.21	12.1	16.11	87.9
16.03	1.679	0.49	6.32	21.79	19.66	-0.098	2.25	11.4	17.41	88.6
16.42	1.720	0.49	6.45	22.24	20.33	-0.086	2.27	11.1	18.07	88.9
16.49	1.727	0.49								
17.57	1.840	0.49								
17.86	1.870	0.49	6.60	22.76	22.86	0.004	2.32	10.2	20.54	89.8
17.99	1.884	0.49								
13.82	1.447	0.39								
13.37	1.400	0.34								
11.55	1.209	0.28								
12.49	1.308	0.28								
14.11	1.478	0.28	8.30	20.00	13.78	-0.311	2.16	15.7	11.62	84.3
12.61	1.320	0.15								
5.83	0.611	0.15								

Table 5.11 Prediction of Mass Transfer Coefficient (Batch Study Data For VOCs)

Rotational Speed, w (rpm)	(rad/s)	Rh (H/R)	KLa 10-3 m/s	KL 10-6 m/s obs.	pred.	error	Chloroform KL1 10-3 1/s	% KL	KLd 10-3 1/s	% KL
1.62	0.170	0.49								
3.50	0.367	0.49								
4.49	0.470	0.49	0.95	3.28	4.02	0.226	1.53	38.2	2.48	61.8
7.76	0.813	0.49	2.01	6.93	7.55	0.089	1.81	24.0	5.74	76.0
10.80	1.131	0.49	2.90	10.00	11.51	0.151	2.00	17.4	9.51	82.6
11.22	1.175	0.49								
11.83	1.239	0.49	3.90	13.45	12.99	-0.034	2.05	15.8	10.94	84.2
11.94	1.250	0.49	3.70	12.76	13.15	0.030	2.06	15.7	11.09	84.3
12.35	1.293	0.49								
12.80	1.340	0.49	4.10	14.14	14.43	0.021	2.10	14.6	12.33	85.4
13.75	1.440	0.49								
15.01	1.572	0.49								
15.24	1.596	0.49	5.60	19.31	18.33	-0.051	2.21	12.1	16.11	87.9
16.03	1.679	0.49	5.90	20.34	19.66	-0.033	2.25	11.4	17.41	88.6
16.42	1.720	0.49	6.30	21.72	20.33	-0.064	2.27	11.1	18.07	88.9
16.49	1.727	0.49								
17.57	1.840	0.49								
17.86	1.870	0.49	6.20	21.38	22.86	0.069	2.32	10.2	20.54	89.8
17.99	1.884	0.49								
13.82	1.447	0.39								
13.37	1.400	0.34								
11.55	1.209	0.28								
12.49	1.308	0.28								
14.11	1.478	0.28								
12.61	1.320	0.15								
5.83	0.611	0.15								

Table 5.12 Values of KL Predicted by Eq. 5.5 (Continuous Flow)

No.			12-3	11-29	11-28	11-26	11-25	11-19	11-12	11-11	11-3	10-30	10-25
w,rpm			12.4	0.82	4.59	1.17	3.99	7.37	5.52	5.58	4.89	1.63	1.63
rad/s			1.3	0.085	0.48	0.12	0.42	0.77	0.58	0.59	0.51	0.17	0.17
Flow,ml/min			386.4	363	380	382	382	382.2	400	237	131.8	132.2	324.1
Tank Tempt., C			17	21	16	21	20	16	16	23	19	20	20
Average			KL 10-6 m/sec (observed)										
CLF			14.637	0.886	4.236	1.334	4.258	7.833	5.510	5.884	4.958	2.525	2.114
TCEN			15.363	0.993	4.357	1.522	4.407	8.304	5.648	6.001	5.093	2.503	2.239
TCE			15.604	1.619	5.065	2.217	4.798	8.685	5.988	6.611	5.309	3.202	2.964
PCE			15.947	1.360	4.793	1.947	4.759	8.737	6.001	6.628	5.347	2.945	2.713
EB			15.978	1.691	5.034	2.352	4.957	8.841	6.207	6.931	5.424	3.339	3.093
Liquid Film													
	Nd	Cd 10-6	KLd 10-6 (m/s)										
CLF	1.53	16.25	14.692	0.226	3.199	0.384	2.608	6.593	4.274	4.387	3.510	0.654	0.654
TCEN	1.53	16.25	14.692	0.226	3.199	0.384	2.608	6.593	4.274	4.387	3.510	0.654	0.654
TCE	1.53	17.25	15.597	0.240	3.396	0.407	2.769	6.999	4.537	4.657	3.726	0.694	0.694
PCE	1.53	17.25	15.597	0.240	3.396	0.407	2.769	6.999	4.537	4.657	3.726	0.694	0.694
EB	1.53	17.15	15.506	0.239	3.377	0.405	2.753	6.958	4.510	4.630	3.705	0.690	0.690
Water Surface													
	Nl	Cl 10-6	KLl 10-6 (m/s)										
CLF	0.22	38.5	2.039	1.119	1.638	1.207	1.591	1.817	1.708	1.714	1.660	1.304	1.304
TCEN	0.22	38.5	2.039	1.119	1.638	1.207	1.591	1.817	1.708	1.714	1.660	1.304	1.304
TCE	0.22	47.0	2.490	1.366	2.000	1.474	1.942	2.219	2.085	2.092	2.026	1.591	1.591
PCE	0.22	47.5	2.516	1.381	2.021	1.490	1.962	2.242	2.107	2.115	2.048	1.608	1.608
EB	0.22	47.5	2.516	1.381	2.021	1.490	1.962	2.242	2.107	2.115	2.048	1.608	1.608
Overall KL (Predicted)			KL 10-6 (m/s)										
CLF			16.732	1.346	4.837	1.591	4.199	8.411	5.981	6.101	5.170	1.957	1.957
TCEN			16.732	1.346	4.837	1.591	4.199	8.411	5.981	6.101	5.170	1.957	1.957
TCE			18.086	1.607	5.396	1.881	4.710	9.218	6.621	6.749	5.753	2.285	2.285
PCE			18.113	1.621	5.417	1.897	4.731	9.241	6.643	6.772	5.774	2.302	2.302
EB			18.022	1.620	5.397	1.895	4.715	9.201	6.617	6.745	5.753	2.298	2.298
Error of Predicted KL													
CLF			0.143	0.519	0.142	0.193	-0.014	0.074	0.086	0.037	0.043	-0.225	-0.074
TCEN			0.089	0.355	0.110	0.046	-0.047	0.013	0.059	0.017	0.015	-0.218	-0.126
TCE			0.159	-0.008	0.065	-0.152	-0.018	0.061	0.106	0.021	0.084	-0.286	-0.229
PCE			0.136	0.192	0.130	-0.026	-0.006	0.058	0.107	0.022	0.080	-0.218	-0.151
EB			0.128	-0.042	0.072	-0.195	-0.049	0.041	0.066	-0.027	0.061	-0.312	-0.257

Table 5.12 Values of KL Predicted by Eq. 5.5 (Continuous Flow)

No.			10-23	10-18	10-15	10-11	10-8	10-6	9-21	9-19	9-18	9-15	9-14
w.rpm			1.63	12.25	11.7	11.15	7.57	12.04	8.08	9.12	9.78	8.28	11.14
rad/s			0.17	1.28	1.23	1.17	0.79	1.26	0.92	0.96	1.02	0.87	1.17
Flow,ml/min			249.6	224	203	328.57	130.5	131	206	210	373.5	252	327
Tank Tempt., C			15	20	20	20	19	20	20	23	18	21	19
Average													
CLF			1.756	15.395	15.021	14.157	8.547	15.724	10.403	9.483	11.475	9.238	16.199
TCEN			2.074	17.472	18.362	15.493	9.349	16.664	11.653	11.483	13.146	9.731	17.535
TCE			3.130	16.628	16.217	15.375	9.412	16.019	10.916	11.807	12.927	10.333	16.562
PCE			2.534	17.347	17.941	15.833	9.178	16.889	11.607	11.868	12.654	10.234	17.881
EB			3.188	18.727	17.910	16.220	9.349	16.992	11.791	12.136	13.837	10.467	17.467
Liquid Film													
	Nd	Cd 10-6											
CLF	1.53	16.25	0.654	14.348	13.499	12.505	6.857	14.006	8.657	9.239	10.137	7.947	12.505
TCEN	1.53	16.25	0.654	14.348	13.499	12.505	6.857	14.006	8.657	9.239	10.137	7.947	12.505
TCE	1.53	17.25	0.694	15.231	14.330	13.275	7.279	14.868	9.190	9.808	10.761	8.436	13.275
PCE	1.53	17.25	0.694	15.231	14.330	13.275	7.279	14.868	9.190	9.808	10.761	8.436	13.275
EB	1.53	17.15	0.690	15.143	14.247	13.198	7.237	14.782	9.136	9.751	10.699	8.388	13.198
Water Surface													
	Nl	Cl 10-6											
CLF	0.22	38.5	1.304	2.032	2.015	1.993	1.828	2.025	1.890	1.908	1.933	1.867	1.993
TCEN	0.22	38.5	1.304	2.032	2.015	1.993	1.828	2.025	1.890	1.908	1.933	1.867	1.993
TCE	0.22	47.0	1.591	2.481	2.460	2.433	2.231	2.473	2.307	2.329	2.360	2.279	2.433
PCE	0.22	47.5	1.608	2.508	2.486	2.458	2.255	2.499	2.332	2.354	2.385	2.303	2.458
EB	0.22	47.5	1.608	2.508	2.486	2.458	2.255	2.499	2.332	2.354	2.385	2.303	2.458
Overall KL (Predicted)													
CLF			1.957	16.380	15.514	14.498	8.685	16.032	10.547	11.147	12.071	9.814	14.498
TCEN			1.957	16.380	15.514	14.498	8.685	16.032	10.547	11.147	12.071	9.814	14.498
TCE			2.285	17.712	16.790	15.707	9.510	17.341	11.497	12.137	13.121	10.716	15.707
PCE			2.302	17.739	16.816	15.733	9.534	17.367	11.521	12.162	13.146	10.740	15.733
EB			2.298	17.650	16.733	15.656	9.492	17.281	11.468	12.105	13.084	10.691	15.656
Error of Predicted KL													
CLF			0.115	0.064	0.033	0.024	0.016	0.020	0.014	0.176	0.052	0.062	-0.105
TCEN			-0.056	-0.062	-0.155	-0.064	-0.071	-0.038	-0.095	-0.029	-0.082	0.009	-0.173
TCE			-0.270	0.065	0.035	0.022	0.010	0.083	0.053	0.028	0.015	0.037	-0.052
PCE			-0.092	0.023	-0.063	-0.006	0.039	0.028	-0.007	0.025	0.039	0.049	-0.120
EB			-0.279	-0.057	-0.066	-0.035	0.015	0.017	-0.027	-0.003	-0.054	0.021	-0.104

Fig 5.11 KL vs Rotational Speed
(Batch Study, Rh=0.49)

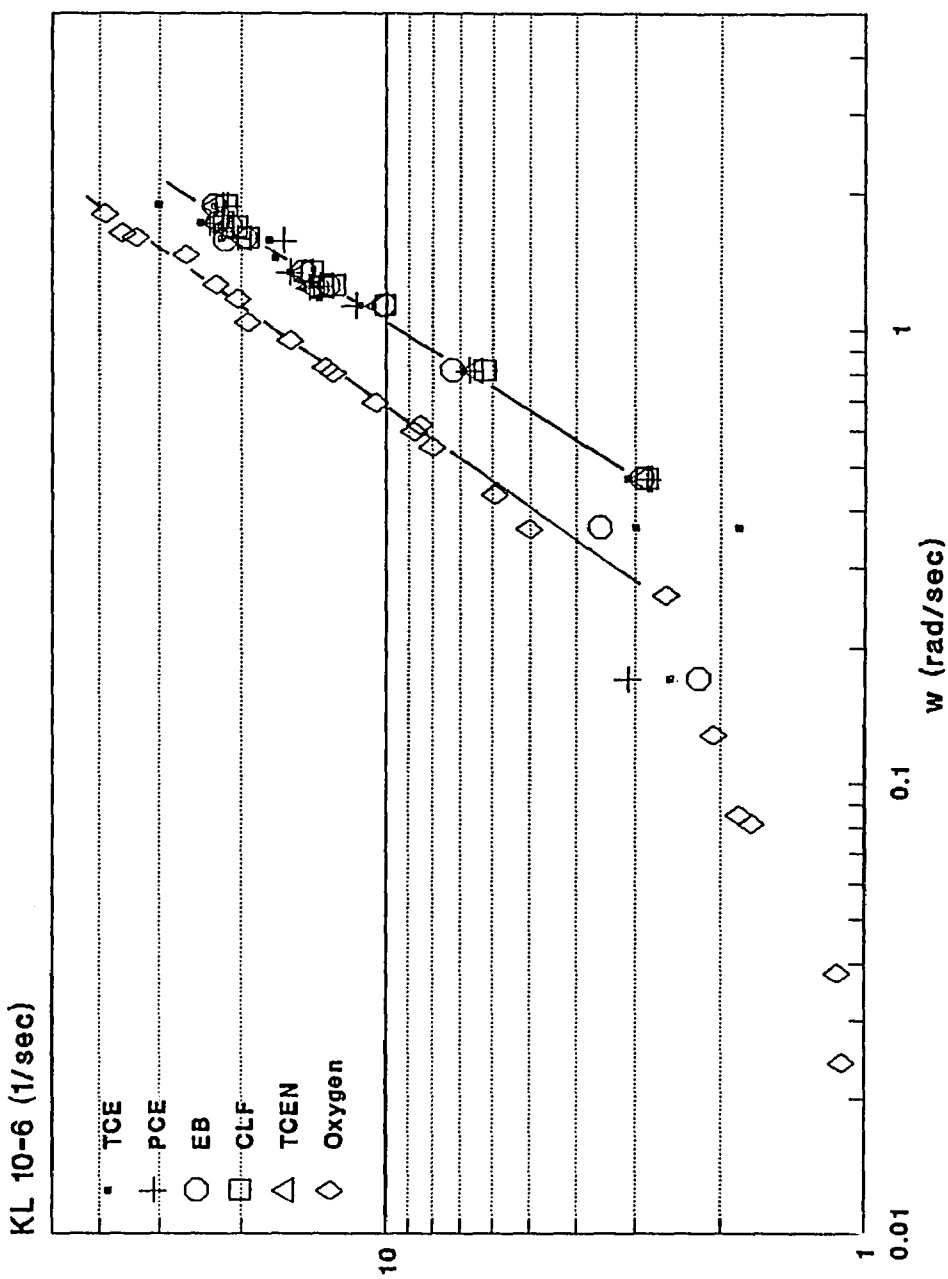


Fig 5.12 Relations between KL and w
(Continuous Flow)

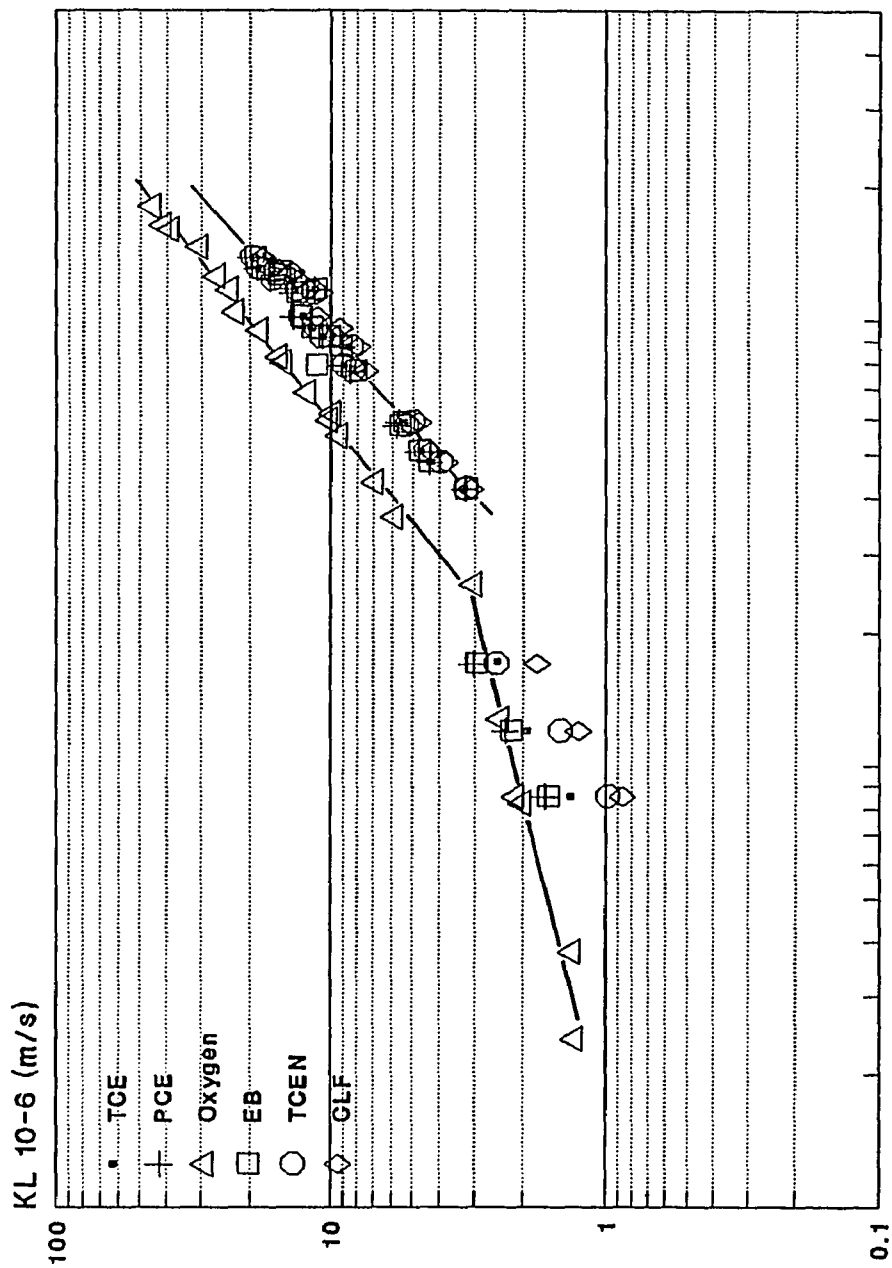
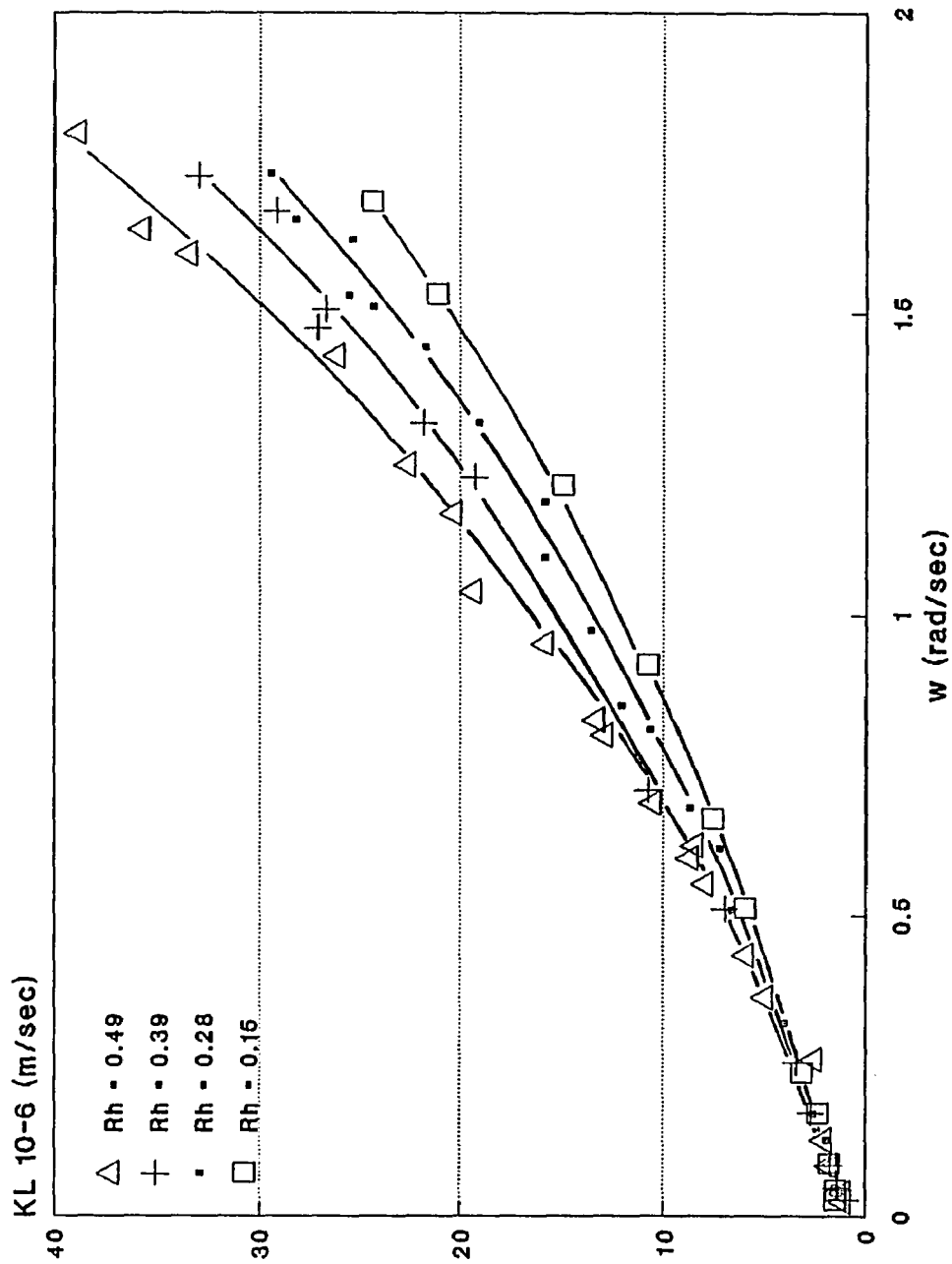


Fig 5.13 Relations between KL, w, Rh
(Oxygen, Batch Study)



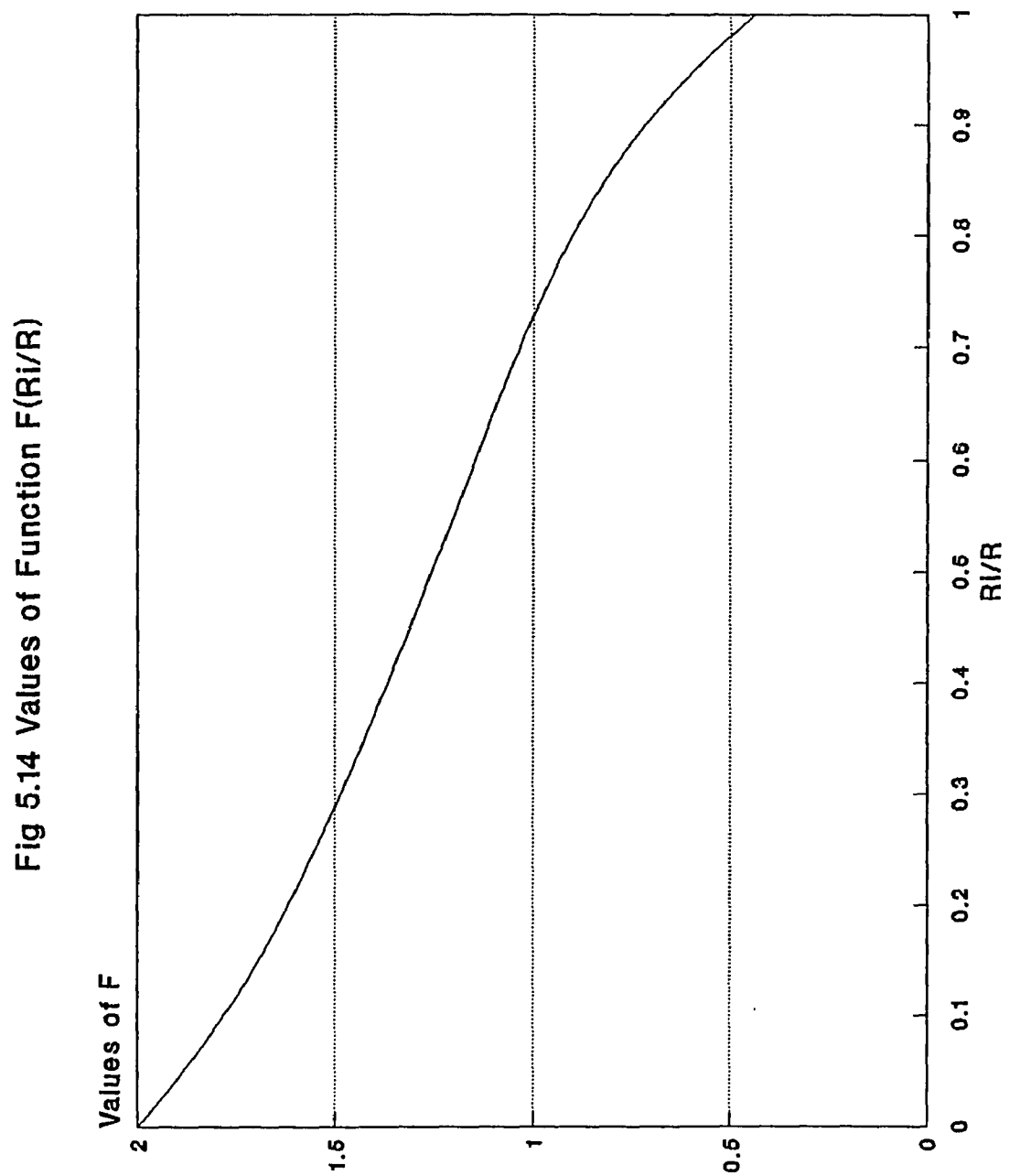


Fig 5.14 Values of Function F(Ri/R)

Fig 5.15 Relation Between C and w
(Batch Study, Rh=0.49)

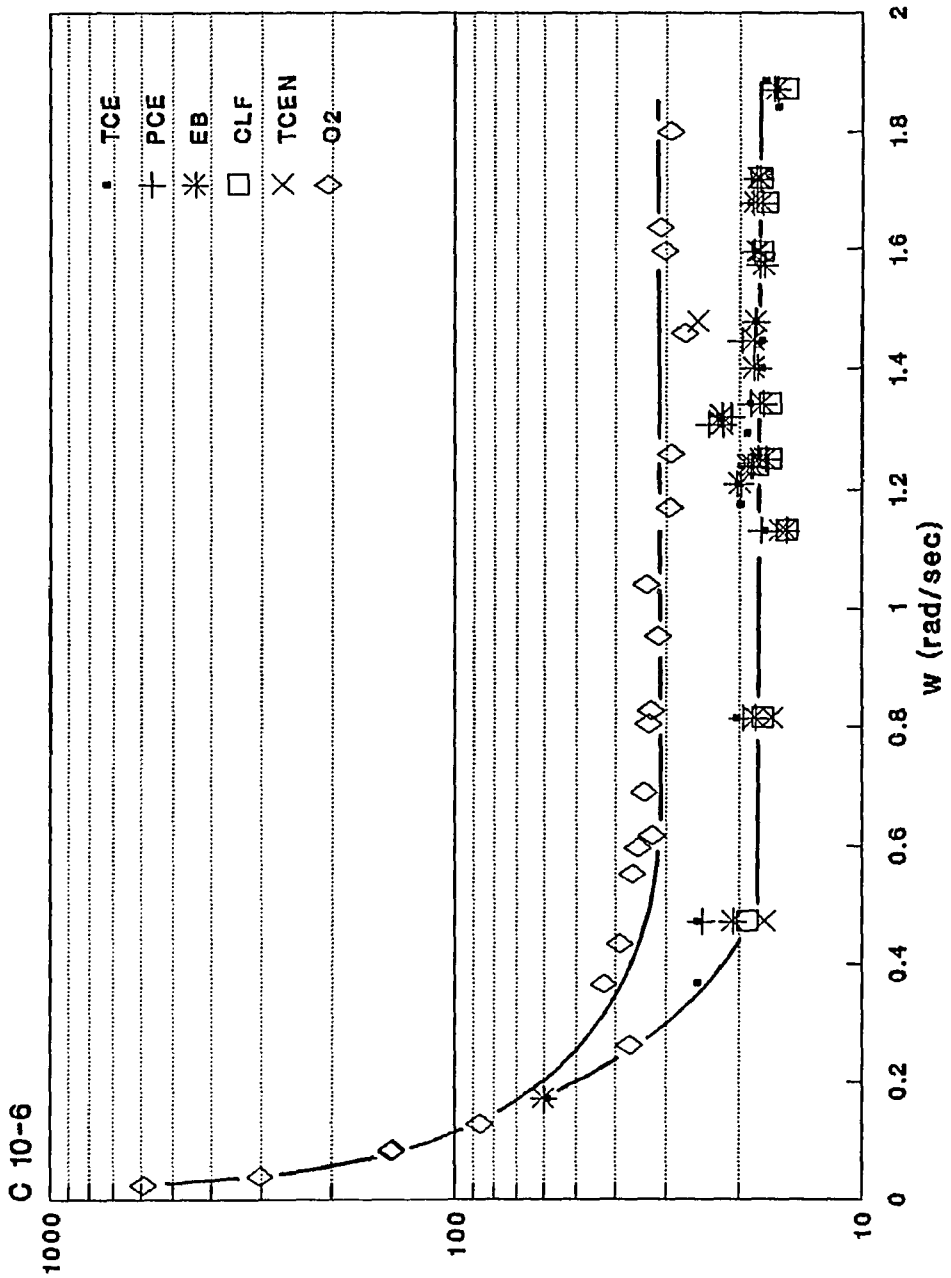


Fig 5.16 Relationships Between C and w
(Continuous Flow Study)

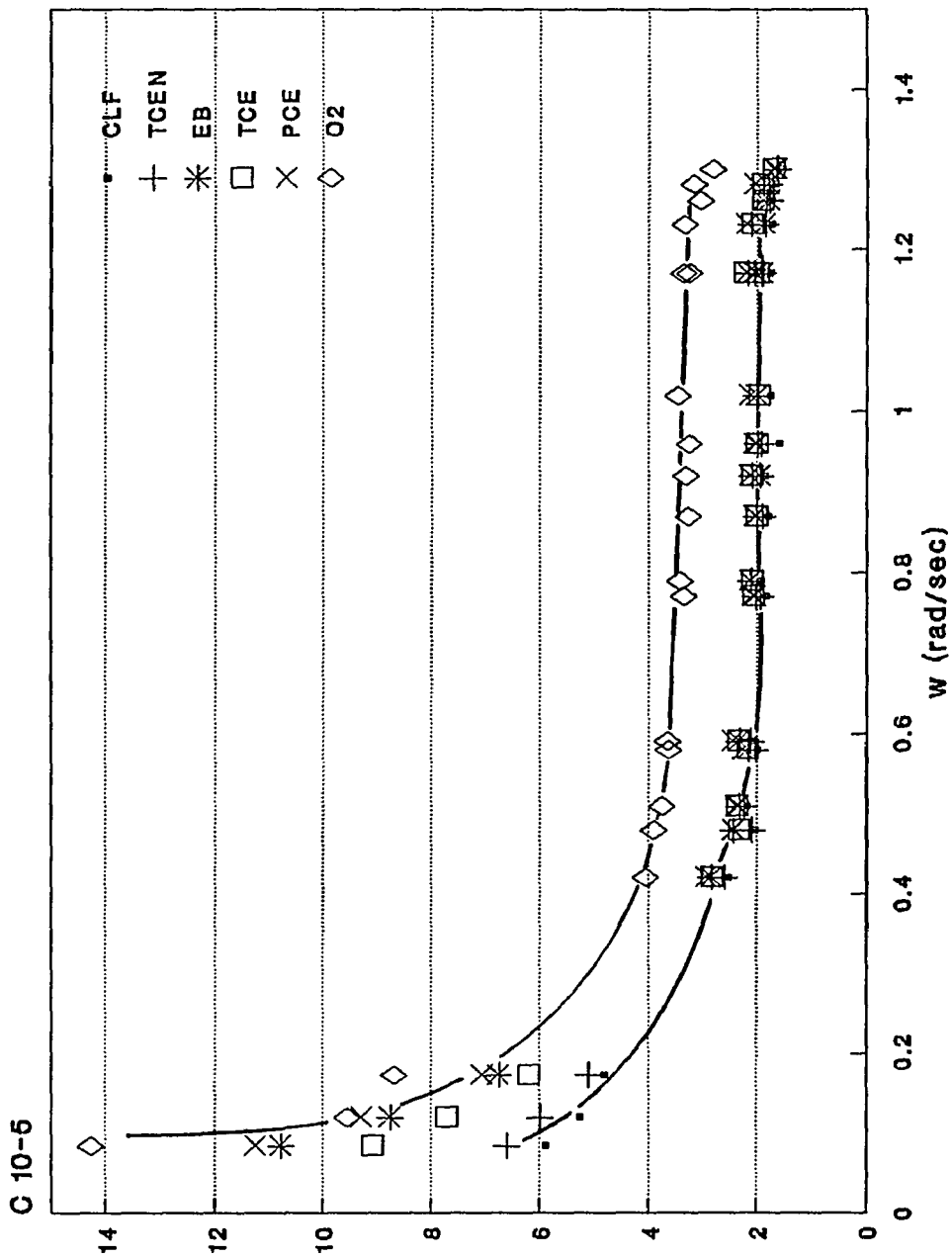


Fig 5.17 Experimentally Determined C
(Batch Study, Oxygen)

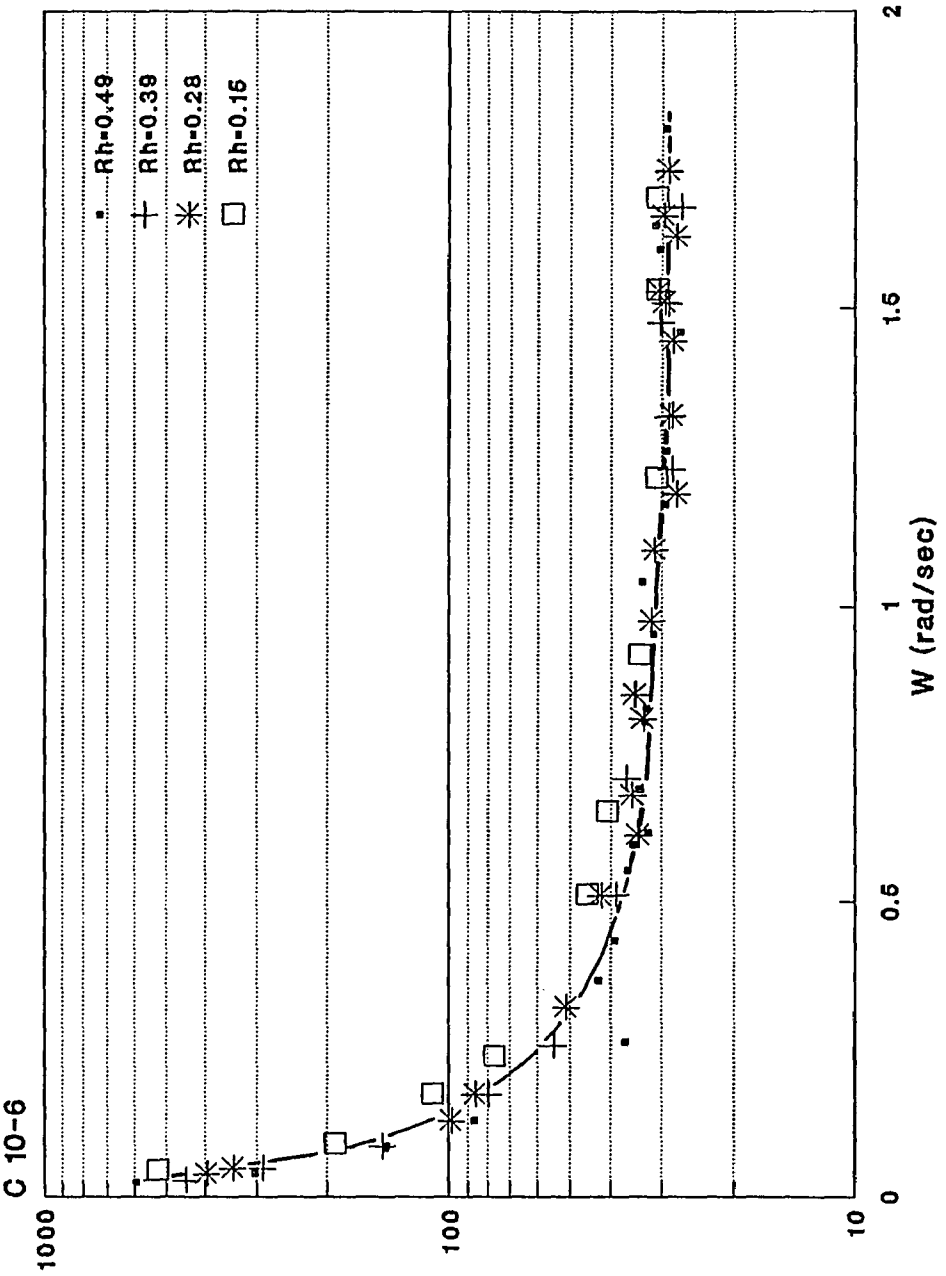


Fig 5.18 Mass Transfer from Water Surface and Disk Surface

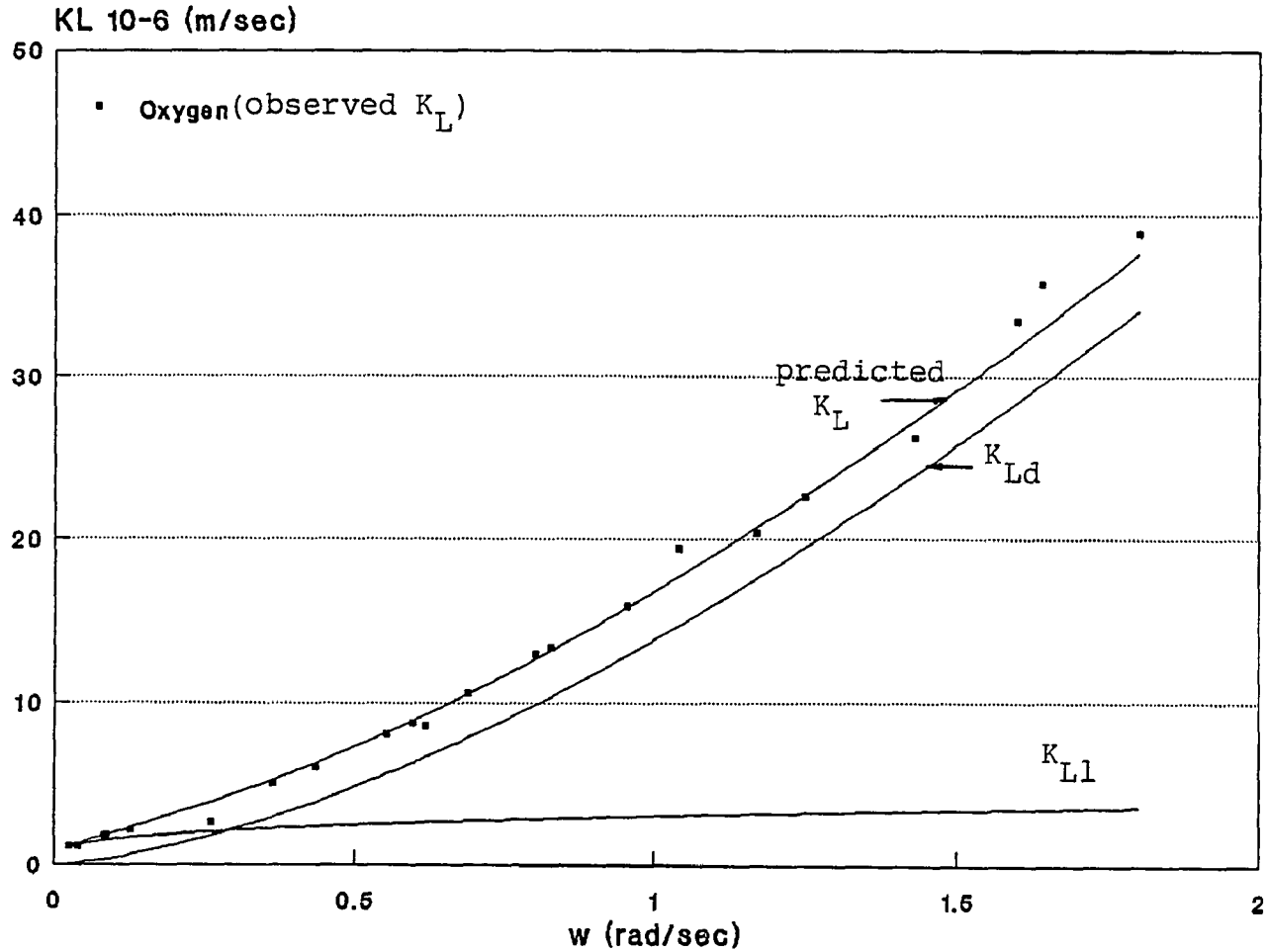
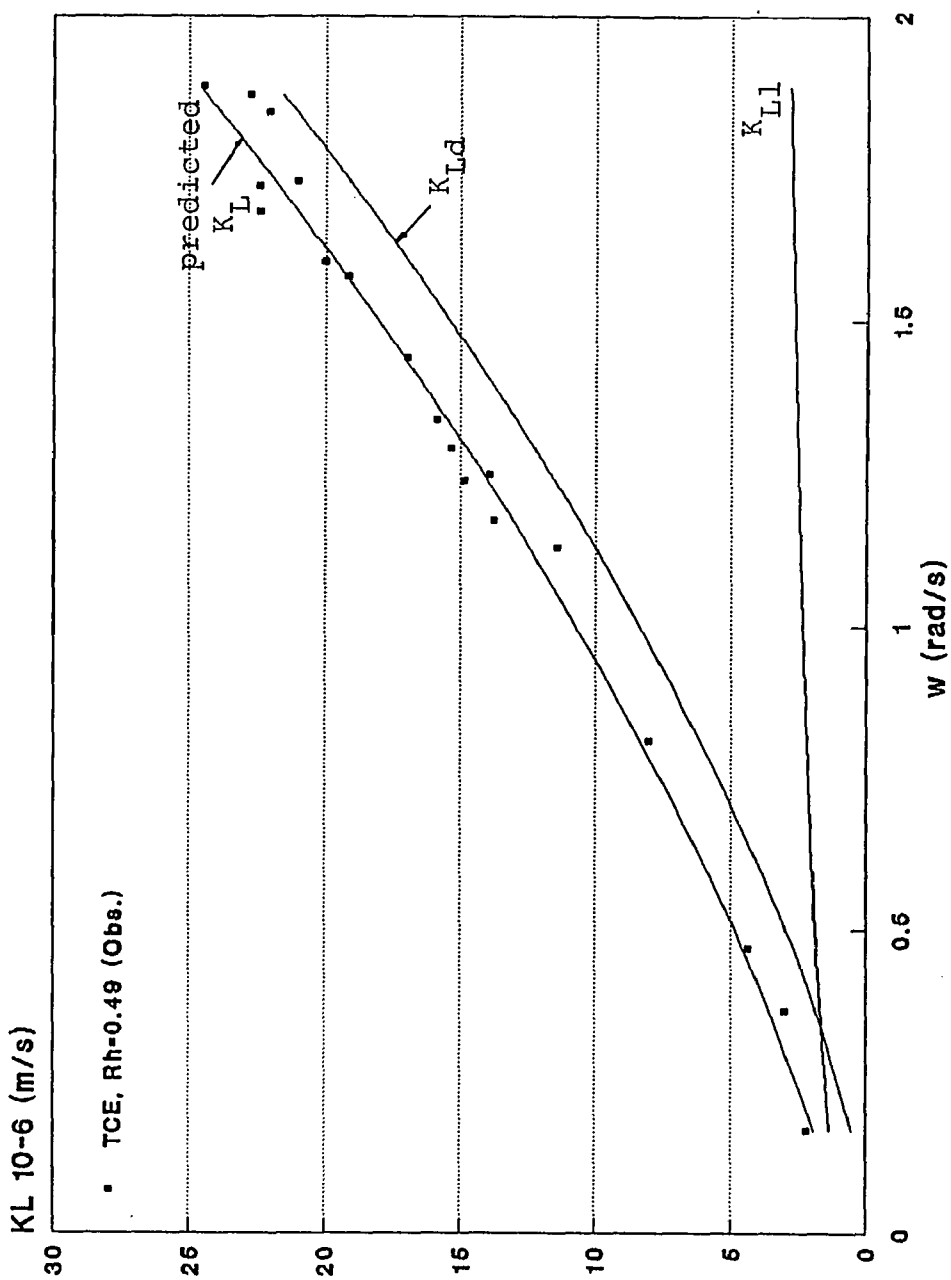


Fig 5.19 Transfer from Water Surface and Disk Surface



5.5 The Effects of Diffusivity and Henry's Constant on Air Stripping of VOCs

For air/water interphase mass transfer, there are resistances to mass transfer in both the gas and liquid phases. The overall mass transfer coefficient can be equated as equation 3.4, in which the total resistance, $R_T=1/K_La$, liquid phase resistance $R_L=1/k_L$, and gas phase resistance, $R_G=1/Hc\ k_Ga$. This simple model was first postulated by Whitman (1923) which assumes that the absorption/desorption rates are controlled only by diffusion on each side of the gas-liquid interface and no resistance exists at the phase boundary. Because of the greater molecular packing density in a liquid system, molecular diffusion of oxygen and nonpolar organic substances in air are roughly 1,000 to 10,000 times faster than in water. When the solute is only slightly soluble, its large Hc makes the gas side resistance negligible. The rate of mass transfer is then controlled by the liquid-phase resistance. In this case, the overall mass transfer coefficient is approximately equal to the liquid-phase transfer coefficient. The gas-phase and liquid-phase diffusivities at 20^o of the selected compounds are listed below (Robert, 1985)

Compounds	D_L x 10^9 m^2/s	D_Gx 10^5 m^2/s	Hc
Oxygen	2.03	2.19	33.60
TCE	0.86	0.85	0.49
PCE	0.79	0.77	1.19
1,1,1 TCEN	0.83	0.77	0.20
CLF	0.94	0.87	0.16
EB	0.90	NA	0.27

Oxygen is a highly volatile component, with high Hc, gas- and liquid phase diffusivities, for which complete liquid phase control has been demonstrated both theoretically and experimentally. For most of the VOCs the liquid-phase resistance is expected to dominate (Smith, 1980, 1981 and Matter-Muller, 1981). However, gas-phase resistance is still expected to exert some degree of influence on the overall mass transfer rate. Literature also shows that more than 90% of the resistance to mass transfer is in the liquid film for highly volatile organic chemicals with Hc values of about 0.001 atm-m^3/mole or higher (Gowda, 1985). Some other studies stated that the liquid-phase resistance dominates for all compounds having a Henry's constant larger than 0.0048 atm-m^3/mol (Mackay, 1975). The physical and chemical characteristics of the compounds studied in this work were listed in Table. All of them satisfy the criterion of liquid-phase control recommended by those researchers. The K_La is assumed to be equal to k_La in this

study.

According to the model developed by Bintanja (1975), if the saturated water film becomes the control mechanism, the mass transfer coefficient will be only the function of film thickness and exposure time (Eq. 3.32). In this case, the values of K_L are the same for all the studied compounds. However, as illustrated in Fig 5.11 and 5.12, the K_L vs w curves for oxygen and VOCs are nearly parallel but not identical. The mass transfer coefficients for VOCs are approximately proportional to that of oxygen. The values of K_La of the five VOCs are close to each other but consistently smaller than that of oxygen. This observation imply that the water film might be highly saturated but not completely saturated. Therefore, Eq. 3.33 may be more close to reality. As discussed previously, the incomplete mixing and enhancement should affect both oxygen and VOCs to the same degree since these two factors are basically influenced by the hydrodynamic properties. The deviations observed between oxygen and VOCs more likely resulted from the different physical and chemical characteristics of oxygen and VOCs in the solution. Since the physical properties of fluids, especially thermal and molecular diffusivities, control the mass transfer at the gas-liquid interface, the mass transfer rate coefficient should be dependent on the molecular diffusivities of the diffusing components. The diffusivities of the VOCs studied

are very close to each other but only about 50% as great as that of oxygen (see Table 4.2). Under similar operating conditions, physical and chemical properties of the fluid, the difference of K_La between oxygen and VOCs may be attributed to their diffusivities. It has been shown in the literature that for a liquid-phase controlled mass transfer process, the ratio of overall mass transfer coefficients of VOCs and of oxygen is independent of the temperature and of the degree of turbulence in the liquid phase (Rathbun, 1978). Since the transfer of oxygen is controlled by liquid phase resistance, it is therefore used as a reference compound and the K_L of each volatile compound was correlated with it.

The experimentally determined diffusivities of the five VOCs were not available in the literature, thus,the diffusivities listed in Table 4.2 was estimated by using the Wilk-Chang correlation (Wilk and Chang, 1955) with the molar volume estimated by means of Schroeder increments (Reid, 1977). The association parameter was assumed to be 2.26 by following the recommendation of Hayduk and Laudie (1974). The accuracy of the estimates of diffusivity is believed to be about 10% (Wilk and Chang, 1955). At present time, the diffusion theories of the liquid have been partly developed, and the existing models do not usually provide as good a prediction of diffusivity as the available empirical equations. Numerous empirical correlations in

addition to Wilk and Chang's equation are available in the literature to estimate the diffusivity in dilute solution. A general comparison of experimental and estimated values of diffusivities in liquids can be found in the literature (Hayduk, 1974).

It is usually found that the diffusivity in liquid is influenced by the concentrations of the diffusing solutes and compositions of the liquid. The concentrations of VOCs or oxygen in this study were as low as less than a few parts per million (ppm). This is also the case for most of the practices of wastewater treatments. Under such low concentrations, it is reasonable to assume that the diffusivity is independent of the concentration. For the same reason, the influence between each component is thought to be negligible as well. The mass transfer process was then considered as a binary diffusion process for each individual diffusing component.

If oxygen is taken as a reference component, a relation between oxygen and VOCs can be written as (Robert, 1983):

$$\frac{(K_L a)_i}{(K_L a)_o} = \left(\frac{D_i}{D_o}\right)^n \tag{5.6}$$

where $(K_L a)_i$, $(K_L a)_o$ are transfer rate constants for VOCs

and oxygen respectively; D_i, D_o are diffusivities of VOCs and oxygen in water. The n, diffusivity exponent, is a constant. The film theory (Whitman, 1923) and the penetration theory (Higbie, 1935) have generally been applied for the study of liquid-side resistance to gas absorption and desorption processes. According to film theory, the mass transfer coefficient is proportional to the diffusivity to the first power; n equals unity. This prediction will usually not agree with the experimental findings from most of the film absorption experiments. The penetration theory predicts that the n equals 0.5. Toor and Marchello (1958) presented a model combining the film and penetration theories, film-penetration model. According to this model, the values of n could vary from 0.5 to unity. The film-penetration theory considers that the entire transfer resistance lies in a laminar surface layer. Surface renewal occurs by eddies which penetrate the surface from the bulk of the liquid. It predicts that the diffusivity exponent varies with changing intensity of turbulence. The value of the exponent could vary from the maximum of unity to the minimum of 0.5, corresponding, respectively, to film diffusion under laminar conditions, and surface renewal mechanism under extreme turbulence. In addition to these three theories, there are still many other models which have been proposed to describe the dependence of the mass transfer coefficient upon

diffusivity.

The values of n in this study were calculated for each VOC at different operating conditions by using the equation

$$n = \ln\left(\frac{(K_L a)_i}{(K_L a)_o}\right) / \ln\left(\frac{D_i}{D_o}\right) \tag{5.7}$$

The average values of the diffusivity exponent n for the individual compounds (w > 4 rpm) were:

Compound	Diffusivity exponent (mean ± std. dev.)
trichloroethylene	0.50 ± 0.045
tetrachloroethylene	0.43 ± 0.032
ethylbenzene	0.54 ± 0.032
chloroform	0.67 ± 0.045
1,1,1 trichloroethane	0.51 ± 0.058

The average values of n, for three compounds (TCE, TCEN, EB) exhibit virtually identical values of 0.49 to 0.52. The n value of chloroform was significantly higher than that of the other compounds, and the n value of tetrachloroethylene was lower. The mean n value of the all five compounds was found to be 0.518. The dependence of the mass transfer rate coefficients on diffusivities was low at lower speed. A comparison between the observed diffusivity exponent and those predicted values of the models is:

Mean value of n	Model Predictions
	Film theory: n = 1.0
	Penetration: n = 0.5
n = 0.53	Film-Penetration:
	0.5 < n < 1.0

According to the prediction of the film-penetration theory n should decrease from unity toward 0.5 with increasing rotational speed, since rotational speed is the most important operational parameter which causes turbulence in the bulk water or in the water film. The values of n from our experiments were plotted in Fig. 5.23 as a function of rotational speeds. Most of the data points scatter around n = 0.53 without showing significant dependence on rotational speed or immersed depth. The film-penetration theory may not be applicable in this case.

Our results favoring the penetration models rather than the film theory is in accordance with many other experiments on gas absorption in packed columns and agitated vessels. Banerjee (1967) reported a proportional relation between K_L and $D^{5/8}$ based on the eddy motion in a wavy laminar film. Yih (1982) proposed K_L was proportional to diffusivity to the power of 2/3. Levich (1962) proposed the dependence of the transfer coefficient on diffusivity

to the power of 0.5. Numerical models and empirical correlations between diffusivity and mass transfer coefficient have been proposed. The diffusivity exponent is frequently observed to lie in the range of 0.5-0.67 (Skelland, 1974) for falling film or stirred tank. For the RDC system, however, not enough data can be found in the literature to confirm the results of this study.

Based on the assumption of liquid phase control, the gas-phase resistance has been ignored in the above data analysis. As seen in Table 5.23, chloroform has the highest diffusivity among the five VOCs, whereas it showed the lowest stripping efficiency. Meanwhile, tetrachloroethylene has the lowest diffusivity but the highest mass transfer rate. The apparent anomalous behaviors of chloroform and tetrachloroethylene may be attributed to the gas-phase resistance. For an open space surface aeration, as used in this study, it is not possible to directly determine the exact values of k_L, and k_G. Since the gas and the liquid-phase diffusivities of the selected VOCs are close to each other, it has been assumed that the values of k_L and k_G are also similar among those compounds. The data was then correlated with Henry's constant in accordance with Eq.3.4. Values of $1/K_La$ were plotted as a function of 1/Hc. Fig. 5.21 shows parts of the results obtained from the continuous flow study as examples.

Linear relationship between these two parameters demonstrated the goodness of fit of the equation. The intercept then gives the mean $1/k_La$, and the slope is the mean $1/k_Ga$. Estimated values of k_La and k_Ga were listed in Table 5.13. A lower value of the ratio k_Ga/k_La or R_L/R_G implies stronger influence of the gas-phase resistance and thus, the less the overall mass transfer coefficient. As indicated in Figure 5.22, both k_Ga and k_La increased with increasing rotational speeds to a power of approximate 1.5. However, the ratio of k_Ga/k_La was relatively independent of w in the region of higher rotational speeds ($w > 4$ rpm). The average value of k_Ga/k_La was calculated to be about 45 in this region.

It is generally assumed that $k_G/k_L = 100$ in natural systems, such as air-sea interface. This value has been generally recommended for evaluating water treatment systems (Kavanaugh, 1980). However, the hydrodynamical conditions are quite different for water treatment systems. The value of the ratio k_G/k_L for the stripping of VOCs during surface aeration was reported to be approximately 25 (Robert, 1983). According to this value, the resistance ratio $R_L/R_T = 0.999$ for oxygen transfer, 0.94 for PCE, 0.90 for TCE, 0.88 for EB, 0.89 for TCEN, and 0.77 for CLF. Although the liquid phase resistance dominates the overall resistance, the gas phase resistance still plays an important role for the transfer of CLF. The values of

k_G/k_L in this study were found to be 45 for the higher speed region which corresponded to R_L/R_T = 0.85 for CLF and 0.97 for PCE. The gas-phase resistance for oxygen transfer accounts for less than 0.1% of the total resistance and can be neglected. The values of k_Ga/k_La were also listed in Table 5.13. For general operational conditions, the liquid phase resistance contributed to more than 90% of the total resistance, but the gas phase resistance had more influence on the stripping of CLF.

In order to correlate the influence of gas phase resistance into the evaluation, Eq. 5.3 is modified to be Eq. 5.5, which is derived by considering $K_La = k_La\ (R_L/R_T)$, and a full liquid-phase controlled mass transfer for oxygen for all operational conditions in this study.

$$\frac{(K_La)_i}{(K_La)_o} = \left(\frac{D_i}{D_o}\right)^n \left(\frac{R_L/R_G}{1 + R_L/R_G}\right) \tag{5.8}$$

The ratio $(K_La)_i/(K_La)_o = 0$, for the compounds of high gas-phase resistance (nonvolatile), and approach a limit equal to $(D_i/D_o)^n$ when R_L goes to infinity (extremely volatile). The values of exponent n were then recalculated by Eq.5.5, and included in Table 5.13. Fig. 5.24 indicates that the recalculated values of n became more consistent to each other than the results from Eq. 5.14. The mean values of n calculated by Eq.5.5 were

Compound	Diffusivity exponent (mean ± std. dev.)
trichloroethylene	0.44 ± 0.048
tetrachloroethylene	0.41 ± 0.034
ethylbenzene	0.45 ± 0.048
chloroform	0.48 ± 0.048
1,1,1 trichloroethane	0.43 ± 0.048

As indicated in Fig.5.11 and 5.12, at low rotational speeds ($w < 4$ rpm), the mass transfer coefficients for the three VOCs, TCE, PCE, EB are about the same as for oxygen. The values of n calculated by Eq. 5.4 show similar results (see Table 5.13). However, the values of n of CLF and TCEN seems to be greater at low rotational speeds, i.e. greater dependence on diffusivity. By considering the resistance of the gas-phase, all five compounds are relatively independent of their diffusivities at low speeds. This observation may imply that film saturation has a dominant role under this condition. According to the model of Zeevalkink, full saturation would occur at about 13 rpm at 30°C. The data from this study indicate that a longer exposure time may be necessary to achieve full water film-saturation in this system. Considering that the film thickness of this system might be thicker than that of a flat disk, a longer exposure time (lower rotational speed) is required. However, since the data obtained in this

region corresponds to low mass transfer rates and higher gas-phase resistance, the reliability of the data is questionable. In addition, both film transfer and water surface transfer are expected to play equally significant roles, making the evaluation more uncertain and more sensitive to experimental errors.

The quantitative estimation of liquid-phase and gas-phase resistance as described above is subject to considerable uncertainty, due to indirectly estimating mean k_G and k_L of the five compounds instead of determining these coefficients for each individual compound. However, the correlation gave a reasonable agreement between predicted values and observed values. The results indicated the dependence of mass transfer rate on diffusivity under normal RDC operations, and further attested the usefulness of the two-resistance approach to interpret the deviation from the diffusivity dependence in mixed resistance mass transfer. Although this approach may only give a crude approximation, this perspective serves as a warning that the assumption of liquid-phase control needs to be evaluated more extensively when extrapolating values of k_G/k_L from other systems to estimate the resistance ratio.

One difficulty in concluding the correctness of

different theories from this study is that the diffusivities of the studied VOCs are only inaccurately known by estimating with the Wilk-Chang correlation, and their values do not vary greatly from each other. However, this is the common case for most VOCs. Since the diffusivities of most VOCs' are close to each other, we may safely estimate the K_La for a VOC from an assumed or experimentally determined value of n and the K_La of oxygen. In our study, n equals 0.52 if Eq. 5.3 is applied, or 0.44 if Eq. 5.5 is applied. Frequently the K_La value for oxygen for a given RBC system is available. The oxygen can then be used as a reference compound to predict the stripping rates of VOCs for a specific RDC system.

5.6 Validity and Application of the Model

With respect to the removal efficiencies of VOCs or the aeration efficiency in a continuous flow RDC system, Eq. 3.7 can be used to design and predict the overall performance of the system. According to this equation, the K_La should be quantitatively evaluated rather than K_L. However, consider the correlations between the experimental data and the operational parameters found in this study and reported in the literature, and the fact that the specific contact area is actually a system dependent constant, the K_L can be used individually to evaluate and design the RDC

system.

The model semiempirically developed in this study provides a more comprehensive correlation between the K_L and those operational parameters. A general form of this model may be expressed as Eq. 5.4. which includes the disk rotational speed, w, the disk radius, R, the specific area of both exposed disk and water surface, a_d and a_l, influence of the immersed depth, F(Ri/R), and two empirical constants. According to the experimental results obtained in this pilot scale study, numerical expression of Eq. 5.4 can be written for each studied compounds as Eq. 5.5. The relation between the predicted and observed values of K_L is shown in Fig 5.25. Shown are the 115 individual experiments, giving a total of about 310 experimental data points. The data encompass a range of approximately 20-fold in the observed K_L. As indicated in the figure, the model gives a reasonable overall fit, as evidenced by the tighter grouping of data around the diagonal line representing perfect agreement. The correlation coefficient, r^2 is equal to 0.95 inferring the goodness of fit of the model.

The values of function F, rotational speeds, and radius of the disk, as included in Eq. 5.4, are dependent on the operations and have known values for each specific

operation. Whereas, the C_l, C_d, n_l, and n_d are not only dependent on the operations but also related to the characteristics of the RDC system. The two coefficients, n_l and n_d, are estimated as 0.33 and 1.53 respectively from the experimental data. In the literature, the values of n_d have been usually found in the range of 1 to 1.6. On the basis of a nearly saturated film-control mechanism, it is safe to apply 1.5 as the value for n_d. The values of n_l estimated in this study are slightly lower than the values reported in the literature. However, the rapid leveling off of K_{Ll} and decrease in significance of the water surface transfer as the rotational speed increases, reduce the possibility that significant errors are induced in estimating the K_L. The deviations of estimated n_l will not cause severe error on the K_L prediction. The predicted valuesof K_L are more sensitive to C_d which is dependenton the properties of the compounds and the RDC systems and has to be determined from the experiments. When data are not available, Eq. 5.5 may be applied as an rough estimates.

For ordinary commercial RDC units, the significance of the water surface transfer is insignificant, compared with the water film transfer from the exposed disk surface (see Table 5.6). For the purpose of VOCs stripping, a higher rotational speed is preferred. Under such operating conditions, Eq. 5.1, 5.2, and the values of C determined from Table 5.8, can be applied to simplify the calculation.

Due to the varieties of configurations and operations of the RDC systems, it is recommended that using oxygen as a reference compounds to estimate the K_L value for each VOC, using Eq. 5.8 or using Eq. 5.6 for liquid phase resistance controlled operations. The oxygen transfer data are usually available for specific RDC systems or can be determined with no difficulty by experiments. With the validity of the relationships of Eq. 3.7 and Eq. 5.1 to 5.8 proven, future applications of those relationships should be valuable in predicting the performance or designing the full-scale RDC systems.

The dimensions of the existing commercial RDC systems are nearly standardized, as shown in Table 2.7 for example. To balance the buoyancy of the water and weight of the media, the depth of submergence is usually kept constant at about 85 to 90% of the disk radius. The operational parameters have been reduced to only the rotational speed and hydraulic loading. The rotational speed is directly related to the energy and maintenance requirements. The hydraulic loading rate is then related to the total capital cost.

To study the relations between the removal efficiency and various operational parameters, a full-scale standard high-density RDC unit with the following specifications was

used,

specific contact area = 243 1/m
total wetted volume = 41.3 m^3/shaft
fraction of immersed disk area = 40% (Rh=0.85)
influent flow rate = 1,500 gpm (0.09464 m^3/sec)

As indicated in Table 5.6, the free water surface contributes only an insignificant portion of the total contact area (less than 0.3%). It is assumed that the overall mass transfer is largely controlled by the liquid film on the disk surface, hence, Eq. 5.1 and Table 5.8 are then used to estimate values of K_L. Based on a 95% removal of TCE, the effects of the rotational speed and hydraulic retention time (in each stage) on the total volume is evaluated by Eq. 3.7 and shown in Fig 5.26a. The required total wetted volume decreases with increasing rotational speed since the values of K_La increase. However, the effect of increasing rotational speed diminishes as w > 8 rpm. For a constant influent flow rate, a lower HRT means the less volume of each stage, and more stages in the system, therefore, the less total volume required.

The power requirement of a full-scale RBC ranging from 1.9 to 3.5 KW per 100,000 square feet of media surface area at 1.6 rpm (Sullivan, 1984). For a specific RDC system, the variation of the power consumption is primarily attributed to the differences in the type and amount of

attached biomass. The energy consumption at 1.6 rpm, without biomass on the disk, was estimated as 1.8 KW per 100,000 sq.ft. of media area by extrapolating Sullivan's data. The power requirement of a RDC system is reported to be proportional to the square of the rotational speed (Sullivan, 1984). The energy requirement for the stripping of VOCs is then evaluated by those relations and illustrated in Fig 5.26b. As indicated in this graph, the increase of energy requirement with increasing rotational speeds becomes more pronounced at higher HRTs of each stage, i.e. less number of stages for a constant influent flow. Considering both Fig 5.26a and 5.26b, a rotational speed in the range of 4 to 8 rpm with smaller HRTs within each stage may give the most economical treatment.

As discussed previously, the staging can increase the overall removal efficiency for a constant reactor volume. As shown in Fig 5.9, the removal efficiency increases markedly as the number of stages is increased from one to two, but the benefit of increasing the number of stages diminishes with increasing stages. Theoretically, the performance of a system of 10 stages in series is scarcely deviated from that of a plug flow system which has the smallest reactor volume. The sizes of commercial RDC units are almost fixed. For the commercial RDC units, the number of stages per shaft can not be increased unlimited since

the specific contact area also decreases with increasing number of stages. A system of four stages per shaft, which is available commercially, may be an appropriate choice for the purpose of VOC stripping. The number of shafts required under each specific operation is also illustrated in Fig 5.26a. The Bare Module Costs (BMC) of a commercial RDC system (Lyco, Inc.) and the packed tower (Ruggiero, 1982), which includes equipment cost as well as direct field material and field labor required for its installation, to remove 95% TCE under various influent flow rates are illustrated in Fig 5.28. Generally, the BMC of the RDC system is lower than of the packed tower at lower influent flow rates or at higher rotational speeds.

Diffused air aeration and packed tower aeration represent the two extremes of the air/water mass transfer operation. The packed tower aeration provides larger gas-liquid contact area, higher hydraulic loading rate but less contact time. The diffused air aeration provides less contact area and lower hydraulic loading rate but longer contact time. Therefore, the diffused aeration is suitable for the practice of gas absorption, but the packed tower aeration is useful for the purpose of gaseous components desorption. RDC is somewhat classified in between. It has been shown in this study that the mass transfer rate constant is relatively independent of the hydraulic loading rate and gas flow rate in an RDC system

which makes the RDC system more flexible for various purposes.

Due to the inherently difference in optimizing various processes for air stripping, it is only possible in this study to use examples as a comparison. Generally, to achieve a specific stripping efficiency in the packed towers, the increase of air-to-water ratio decreases the tower height but increases the tower diameter or gas pressure drop. As the gas pressure drop increases, smaller tower volumes are obtained. The total brake power requirements are determined from the gas pressure drop, the air-to-water ratio, and the type of packing. For a 1,500 gpm influent flow rate and 95% TCE removal efficiency, the relations between tower volume, total brake power and air-to-water ratio are presented by (Hand, 1986).

A comparison between RDC and packed tower shows somewhat lower energy requirements for the packed tower than for the RDC system. The comparison of energy probably is biased, since the energy calculated for RDC system was extrapolated from the systems with biomass attatching on the disk surface. The energy comsmption is certainly lower for clean disks, besides the surface configurations also influence the energy requirement. The presented energy requirements should not be construed as

absolute values but rather as indications of the relative energy comsumptions under different operations for both processes. The RDC also entails a greater volume than the packed tower, whereas, the construction cost per unit volume of RDC is somewhat lower than the pack tower system, since no extra support structure is required for RDC. The approach to optimizing packed tower design for VOCs removal is difficult in general because of the large number of combinations of parameters involved. For the system and operations reported by Hand (1986) the design parameters was

Air-to-water ratio = 30,

Tower pressure drop = 0.06 in H_2O/ft,

TCE removal = 95%,

Packing type = 3-in plastic saddles,

Water flow rate = 1500 gpm (0.09464 cu m/s),

Tower diameter = 8 ft,

Tower height = 24.5 ft,

Total capital cost = $120,107.00,

Total operational energy consumption = 34.2 KW, for which

blower consumption = 9.2 KW,
pump power consumption = 25 KW.

According to Fig 5.26a and Fig 5.26b, to achieve the same degree of removal, two high density RDC units (four stages in each) with concrete troughs are required. The Bare Module Cost (BMC) for RDC unit with concret trough, as

estimated by Lyco Co., will be only $85,000. The required rotational speed will be 5 rpm which may be the maximal rotational speed of existing RDC units. The energy consumption is estimated by Fig 5.26b as 42.3 KW. The operating and maintenance labor cost is likely to be lower for RDC, especially where there is a risk of clogging by inorganic precipitation, biomass growth, and various solids and particles. The RDC process is simple to install and operate, a great deal of pertinent experience existing with oxygen transfer applications, and the energy requirements are not substantially higher than the packed tower air stripping. In addition, the RDC units are transportable, and will be suitable for some emergent or temporary applications. Therefore, this process may be an alternative for air stripping when the application of the packed tower is limited, and certainly it deserved further evaluation and study for the VOCs control.

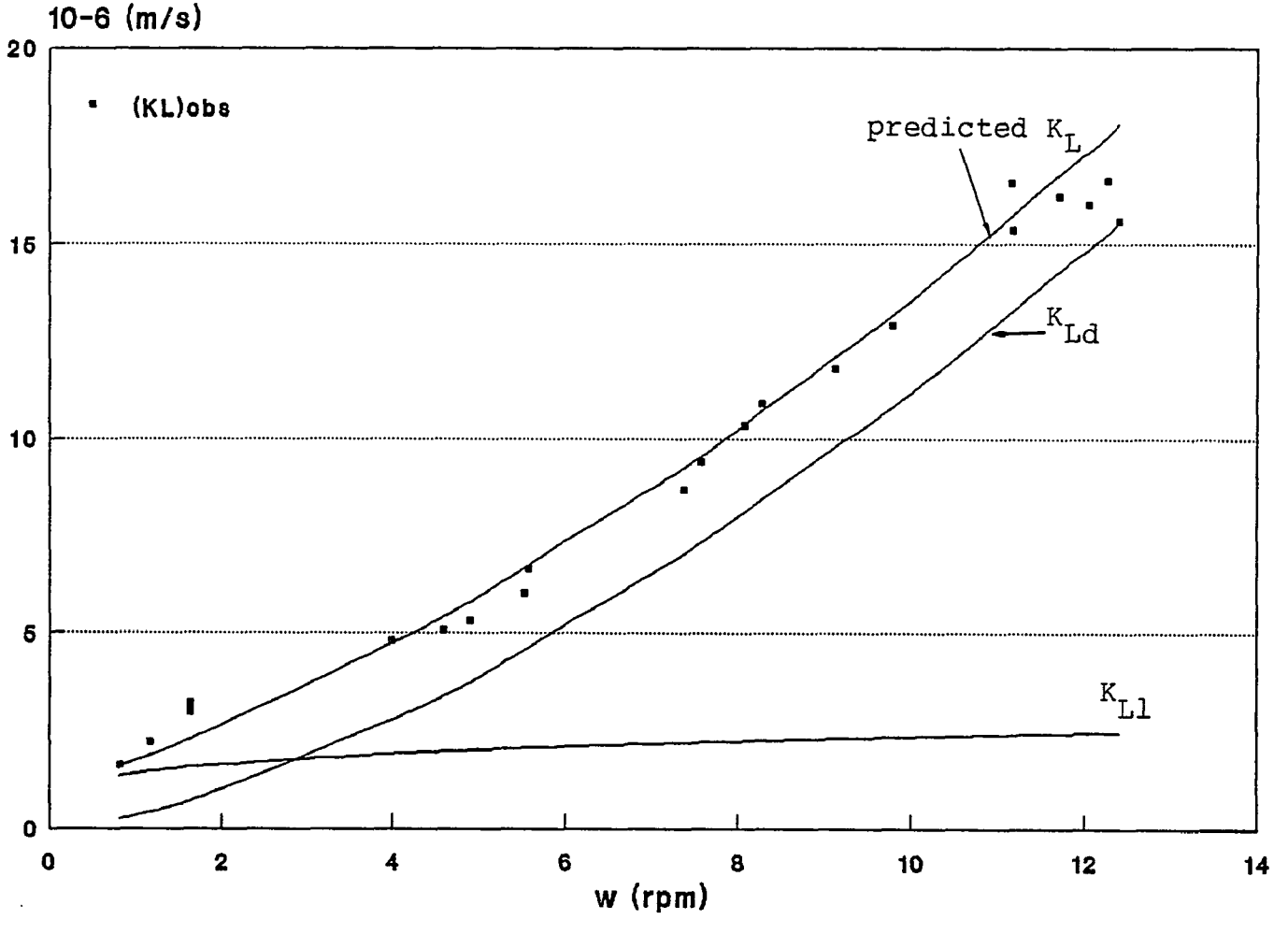
Table 5.13 Calculation of kLa and kGa for VOCs in Continuous Flow 4-Stage RDC System (Calculated by 1st to 3rd stages)

NO.			12-3	11-28	11-25	11-19	11-12	11-11	11-3	10-18	10-6	10-15	10-11
w,rpm			12.4	4.59	3.99	7.37	5.52	5.58	4.89	12.25	12.04	11.7	11.15
rad/s			1.3	0.48	0.42	0.77	0.58	0.59	0.51	1.28	1.26	1.23	1.17
Flow,ml/min			386	380	382	382	400	237	132	224	131	203	329
Water Tempt., C			17	16	20	16	16	23	19	20	20	20	20
KLa Calculation (Average)													
Compounds	Hc	D 10-9,m2/s	KLa (1/sec)										
CLF	0.13	0.94	0.00423	0.00122	0.00123	0.00226	0.00159	0.00170	0.00143	0.00445	0.00454	0.00434	0.00409
TCEN	0.30	0.83	0.00444	0.00126	0.00127	0.00240	0.00163	0.00173	0.00147	0.00505	0.00482	0.00531	0.00448
EB	0.27	0.90	0.00451	0.00146	0.00139	0.00251	0.00173	0.00191	0.00153	0.00481	0.00463	0.00469	0.00444
TCE	0.41	0.86	0.00461	0.00139	0.00138	0.00253	0.00173	0.00192	0.00155	0.00501	0.00488	0.00518	0.00458
PCE	0.82	0.79	0.00462	0.00145	0.00143	0.00255	0.00179	0.00200	0.00157	0.00541	0.00491	0.00546	0.00469
O2	32.1	2.29	0.00769	0.00236	0.00199	0.00415	0.00291	0.00301	0.00248	0.00850	0.00792	0.00839	0.00761
1/(kLa*1000)			0.2088	0.6695	0.6750	0.2079	0.5448	0.4818	0.6237	0.1827	0.2014	0.1784	0.2079
1/(kGa*1000)			0.0040	0.0242	0.0207	0.0047	0.0128	0.0168	0.0110	0.0057	0.0026	0.0064	0.0047
kLa (1/sec)			0.0048	0.0015	0.0015	0.0048	0.0018	0.0021	0.0016	0.0055	0.0050	0.0056	0.0048
kGa (1/sec)			0.2523	0.0414	0.0483	0.2110	0.0781	0.0595	0.0913	0.1761	0.3899	0.1555	0.2109
kGa/kLa			52.6874	27.7226	32.6087	43.8608	42.5293	28.6786	56.9589	32.1712	78.5185	27.7449	43.8515
CompoundsRL/RG for VOCs													
CLF			6.85	3.60	4.24	5.70	5.53	3.73	7.40	4.18	10.21	3.61	5.70
TCEN			15.81	8.32	9.78	13.16	12.76	8.60	17.09	9.65	23.56	8.32	13.16
EB			14.23	7.49	8.80	11.84	11.48	7.74	15.38	8.69	21.20	7.49	11.84
TCE			21.60	11.37	13.37	17.98	17.44	11.76	23.35	13.19	32.19	11.38	17.98
PCE			43.20	22.73	26.74	35.97	34.87	23.52	46.71	26.38	64.39	22.75	35.96
Compounds(RL/RG) /(1+RL/RG) for VOCs													
CLF			0.8726	0.7828	0.8091	0.8508	0.8468	0.7885	0.8810	0.8070	0.9108	0.7829	0.8508
TCEN			0.9405	0.8927	0.9073	0.9294	0.9273	0.8959	0.9447	0.9061	0.9593	0.8927	0.9294
EB			0.9343	0.8821	0.8980	0.9221	0.9199	0.8856	0.9389	0.8968	0.9550	0.8822	0.9221
TCE			0.9558	0.9191	0.9304	0.9473	0.9458	0.9216	0.9589	0.9295	0.9699	0.9192	0.9473
PCE			0.9774	0.9579	0.9639	0.9729	0.9721	0.9592	0.9790	0.9635	0.9847	0.9579	0.9729
Compounds Calculation of n													
CLF	(liquid film)		0.67	0.74	0.54	0.68	0.68	0.64	0.62	0.73	0.62	0.74	0.70
TCEN			0.54	0.62	0.44	0.54	0.57	0.54	0.51	0.51	0.49	0.45	0.52
EB			0.57	0.51	0.39	0.54	0.56	0.49	0.51	0.61	0.57	0.62	0.58
TCE			0.52	0.54	0.38	0.51	0.53	0.46	0.48	0.54	0.49	0.49	0.52
PCE			0.48	0.45	0.31	0.46	0.45	0.38	0.43	0.42	0.45	0.40	0.46
Compounds Calculation of n													
CLF	(two film)		0.52	0.46	0.30	0.50	0.49	0.37	0.47	0.49	0.52	0.47	0.52
TCEN			0.48	0.51	0.34	0.47	0.50	0.43	0.46	0.42	0.45	0.34	0.45
EB			0.50	0.38	0.27	0.45	0.47	0.36	0.45	0.49	0.53	0.49	0.49
TCE			0.48	0.46	0.30	0.45	0.47	0.38	0.44	0.46	0.46	0.41	0.46
PCE			0.46	0.41	0.27	0.43	0.43	0.34	0.41	0.39	0.43	0.36	0.43

Table 5.13 Calculation of kLa and kGa for VOCs in Continuous Flow 4-Stage RDC System (Calculated by 1st to 3rd stages)

NO.			10-8	9-21	9-19	9-18	9-15	9-14	10-30	10-25	10-23	11-29	11-26
w,rpm			7.57	8.08	9.12	9.78	8.28	11.14	1.635	1.632	1.631	0.82	1.17
rad/s			0.79	0.92	0.96	1.02	0.87	1.17	0.171	0.172	0.17	0.085	0.12
Flow,ml/min			131	206	210	374	252	327	132	324	250	363	382
Water Tempt., C			19	20	23	18	21	19	20	20	15	21	21
KLa Calculation (Average)													
Compounds	Hc	D 10-9,m2/s											
CLF	0.13	0.94	0.00247	0.00301	0.00274	0.00332	0.00267	0.00468	0.00073	0.00061	0.00051	0.00026	0.00039
TCEN	0.30	0.83	0.00270	0.00337	0.00332	0.00380	0.00281	0.00507	0.00072	0.00065	0.00060	0.00029	0.00044
EB	0.27	0.90	0.00272	0.00315	0.00341	0.00374	0.00299	0.00479	0.00093	0.00086	0.00090	0.00047	0.00064
TCE	0.41	0.86	0.00265	0.00335	0.00343	0.00366	0.00296	0.00517	0.00085	0.00078	0.00073	0.00039	0.00056
PCE	0.82	0.79	0.00270	0.00341	0.00351	0.00400	0.00302	0.00505	0.00097	0.00089	0.00092	0.00049	0.00068
O2	32.1	2.29	0.00439	0.00538	0.00562	0.00658	0.00488	0.00784	0.00105	0.00110	0.00095	0.00062	0.00070
1/(kLa*1000)			0.3492	0.2872	0.2643	0.2475	0.3199	0.1932	1.0150	1.0580	0.9909	1.7960	1.3250
1/(kGa*1000)			0.0039	0.0057	0.0120	0.0066	0.0076	0.0026	0.0638	0.0949	0.1501	0.3387	0.1990
kLa (1/sec)			0.0029	0.0035	0.0038	0.0040	0.0031	0.0052	0.0010	0.0009	0.0010	0.0006	0.0008
kGa (1/sec)			0.2583	0.1748	0.0833	0.1509	0.1312	0.3880	0.0157	0.0105	0.0067	0.0030	0.0050
kGa/kLa			90.1860	50.2010	22.0067	37.3416	41.9761	74.9709	15.9091	11.1462	6.6016	5.3026	6.6583
CompoundsRL/RG for VOCs													
CLF			11.72	6.53	2.86	4.85	5.46	9.75	2.07	1.45	0.86	0.69	0.87
TCEN			27.06	15.06	6.60	11.20	12.59	22.49	4.77	3.34	1.98	1.59	2.00
EB			24.35	13.55	5.94	10.08	[illegible]3	20.24	4.30	3.01	1.78	1.43	1.80
TCE			36.98	20.58	9.02	15.31		30.74	6.52	4.57	2.71	2.17	2.73
PCE			73.95	41.16	18.05	30.62	34.[illegible]	61.48	13.05	9.14	5.41	4.35	5.46
Compounds(RL/RG) /(1+RL/RG)													
CLF			0.9214	0.8671	0.7410	0.8292	0.8451	0.9069	0.6741	0.5917	0.4618	0.4081	0.4640
TCEN			0.9644	0.9377	0.8685	0.9180	0.9264	0.9574	0.8268	0.7698	0.6645	0.6140	0.6664
EB			0.9606	0.9313	0.8559	0.9098	0.9189	0.9529	0.8112	0.7506	0.6406	0.5888	0.6426
TCE			0.9737	0.9537	0.9002	0.9387	0.9451	0.9685	0.8671	0.8205	0.7302	0.6849	0.7319
PCE			0.9867	0.9763	0.9475	0.9684	0.9718	0.9840	0.9288	0.9014	0.8441	0.8130	0.8452
Compounds	Calculation of n												
CLF	(liquid film)		0.65	0.65	0.81	0.77	0.68	0.58	0.41	0.66	0.70	0.99	0.67
TCEN			0.48	0.46	0.52	0.54	0.54	0.43	0.37	0.52	0.45	0.76	0.46
EB			0.51	0.57	0.53	0.61	0.53	0.53	0.14	0.27	0.05	0.30	0.09
TCE			0.51	0.48	0.50	0.60	0.51	0.43	0.21	0.35	0.27	0.47	0.22
PCE			0.46	0.43	0.44	0.47	0.45	0.41	0.08	0.20	0.03	0.22	0.03
Compounds	Calculation of n												
CLF	(two film)		0.55	0.49	0.47	0.56	0.49	0.47	-0.03	0.07	-0.16	-0.01	-0.19
TCEN			0.44	0.40	0.38	0.46	0.47	0.39	0.18	0.26	0.05	0.28	0.06
EB			0.47	0.50	0.37	0.50	0.44	0.48	-0.09	-0.04	-0.42	-0.27	-0.38
TCE			0.49	0.43	0.40	0.54	0.45	0.39	0.07	0.14	-0.06	0.08	-0.10
PCE			0.44	0.41	0.39	0.44	0.42	0.40	0.01	0.10	-0.13	0.03	-0.13

Fig 5.20 Stripping of TCE from Water Surface and Disk surface



*continuous flow

Fig 5.21 Relation between KLa and Hc
(continuous flow study)

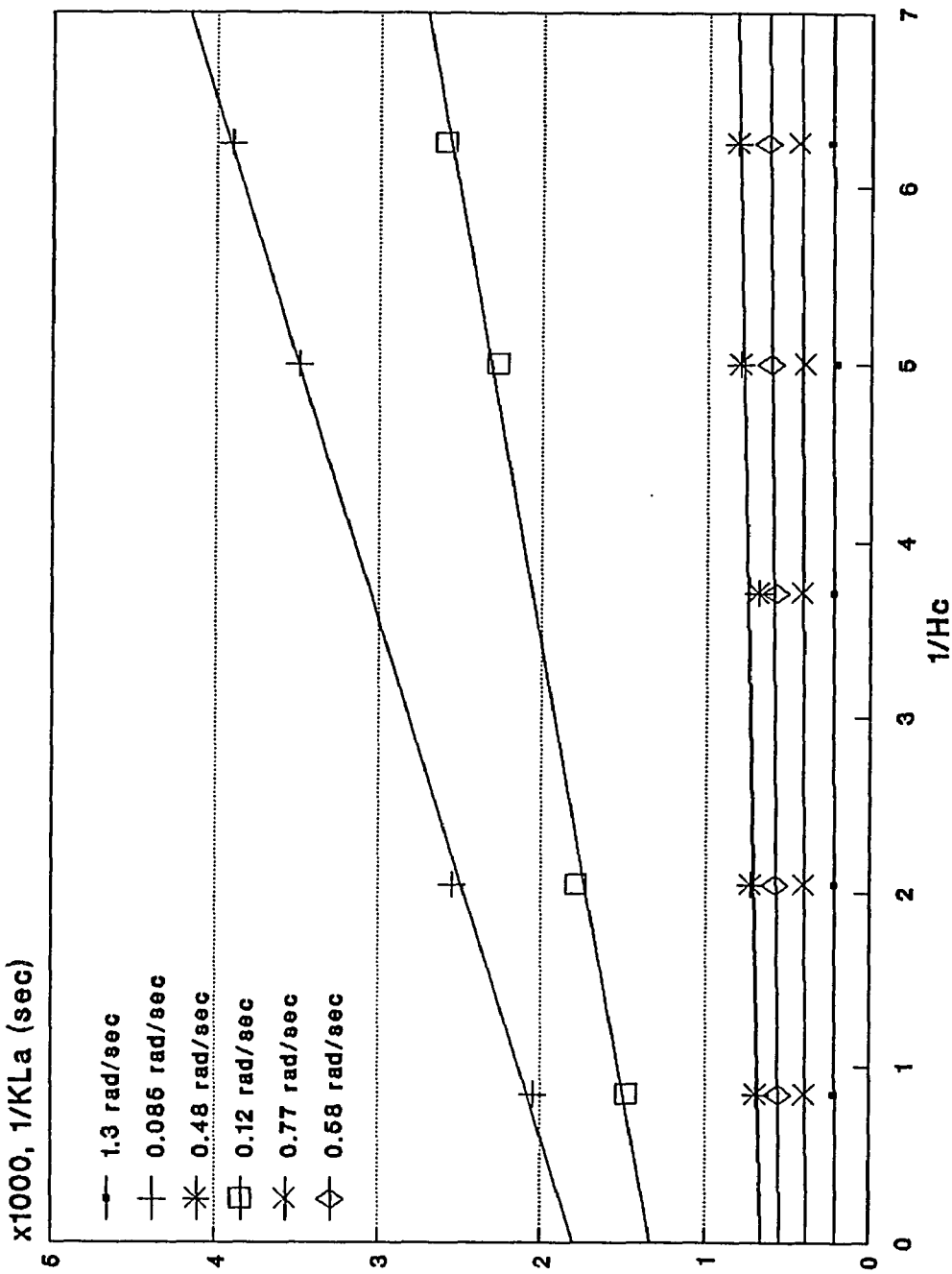


Fig 5.22 kGa, kLa and Rotational Speed

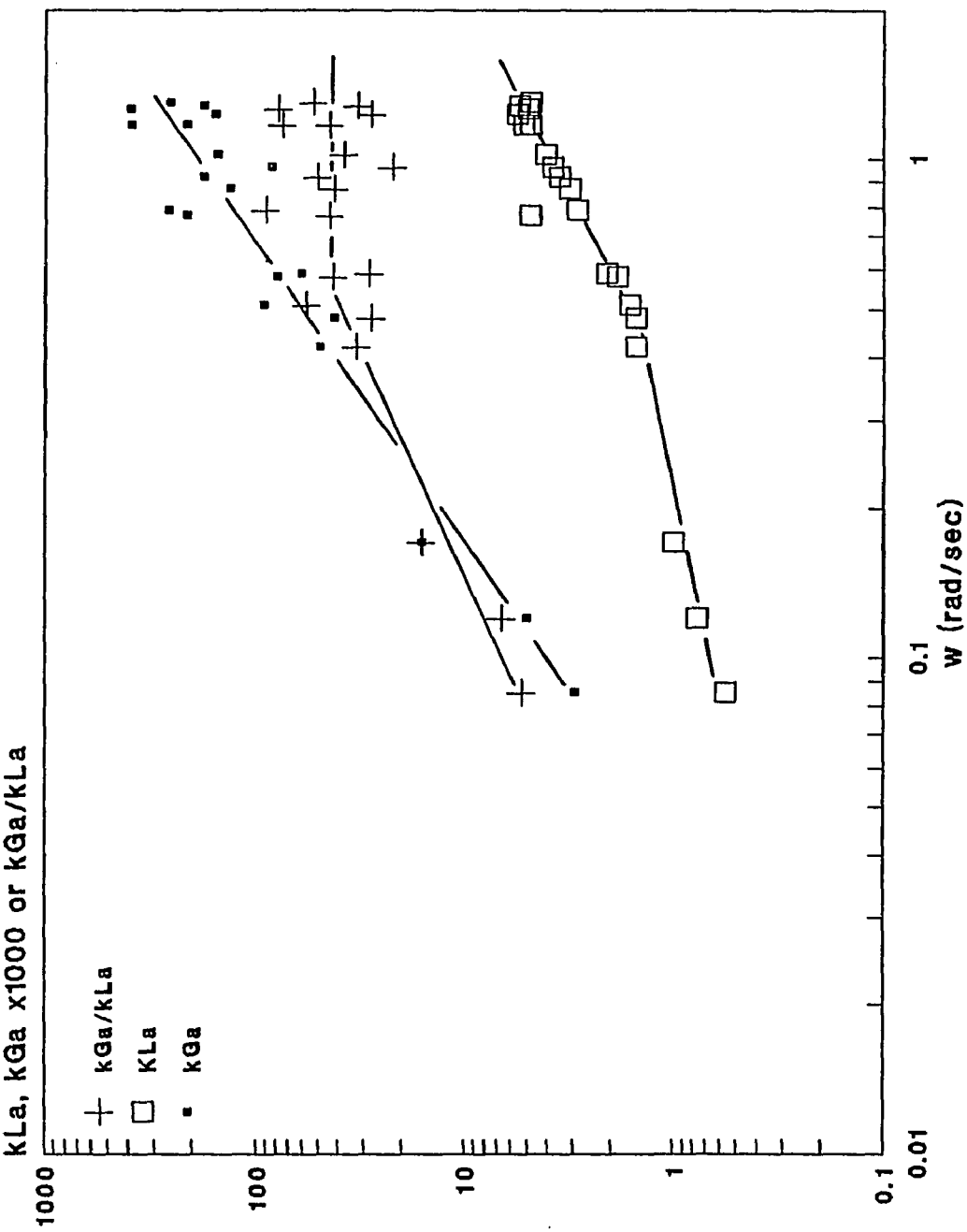
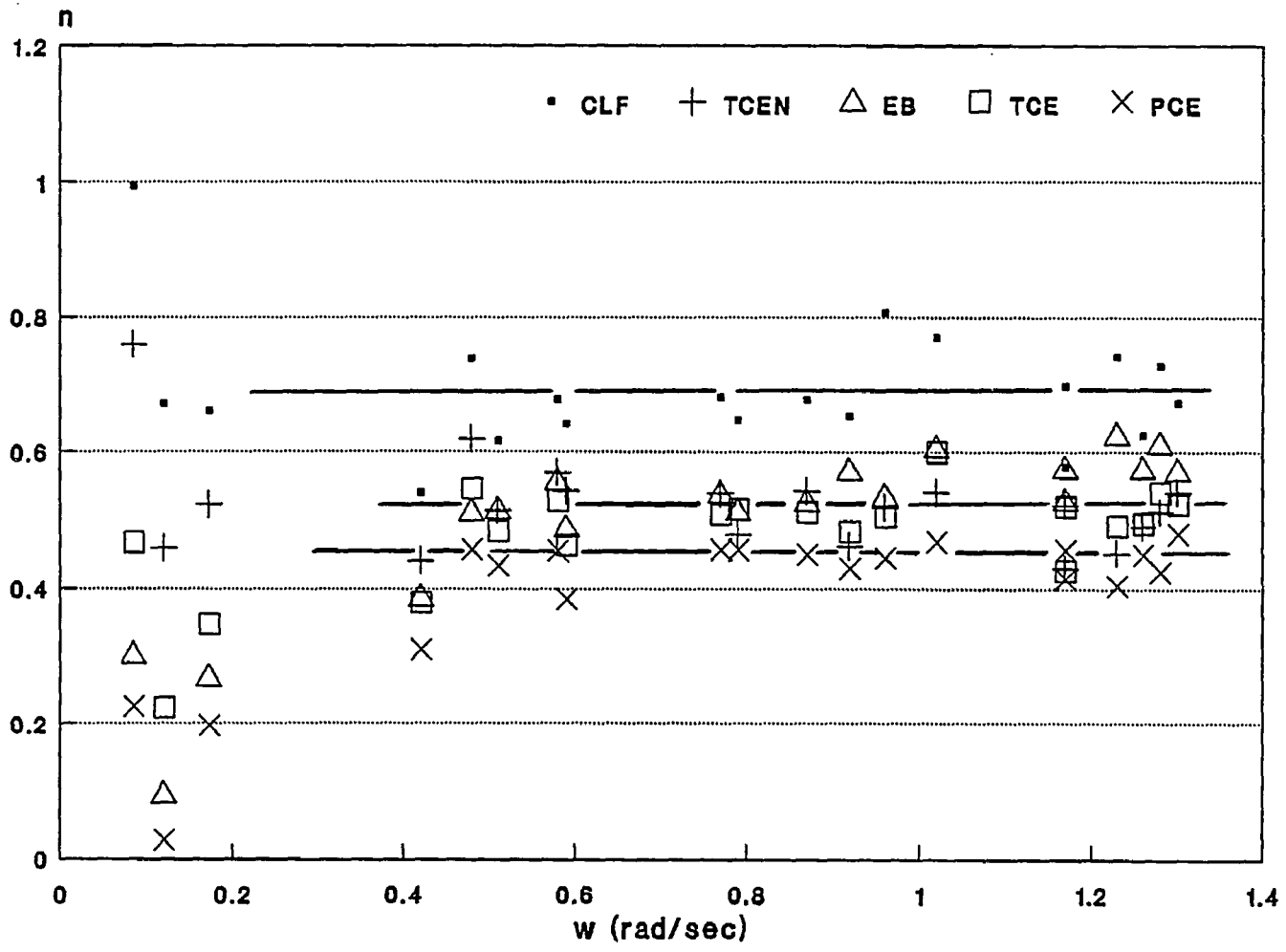


Fig 5.23 Values of n at Different w
(without RG correction)



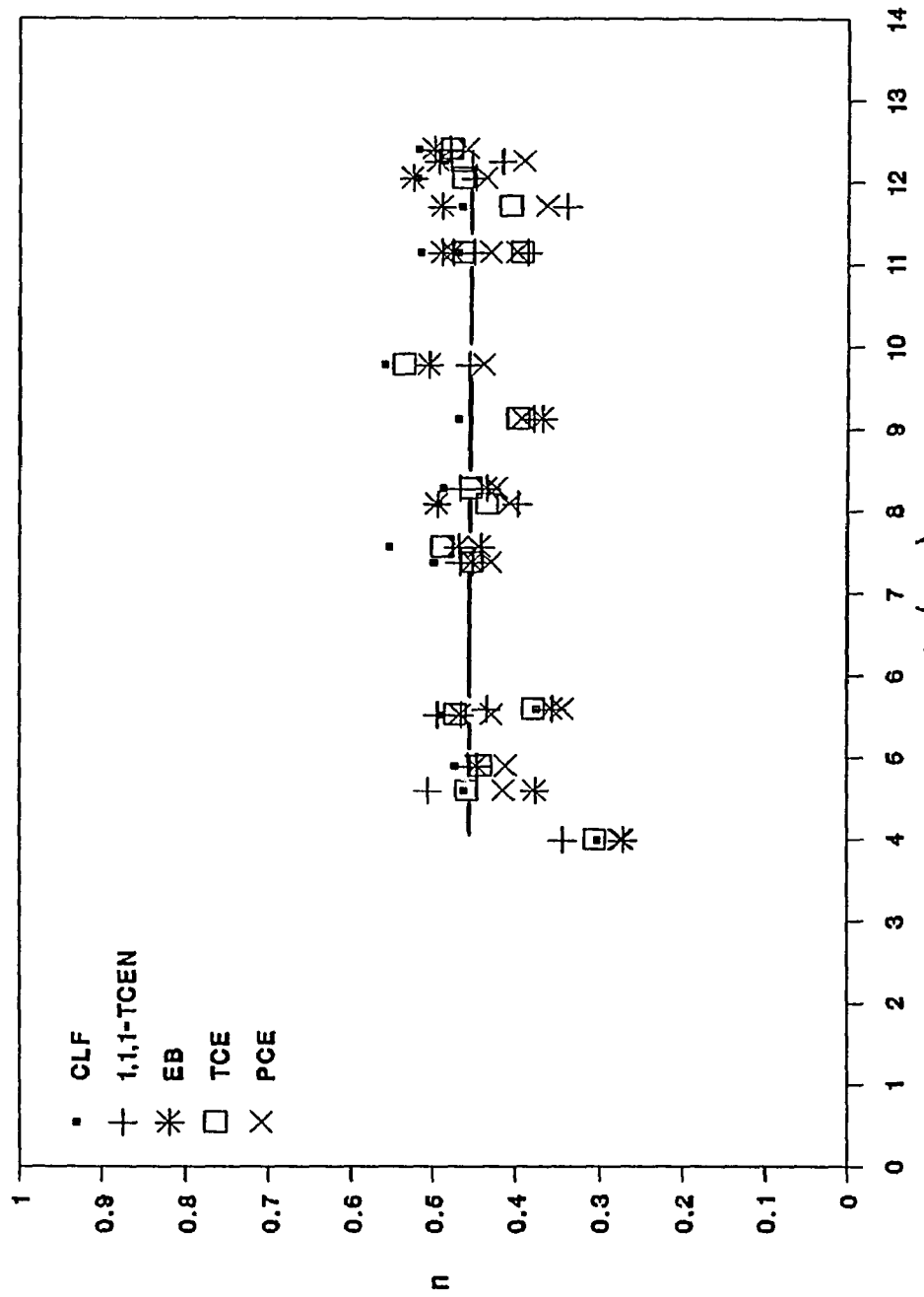


Fig. 5.24 Values of n & Rotational Speed
(with gas-phase resistance correction)

Fig 5.25 Comparison between Observed and Predicted Values of KL

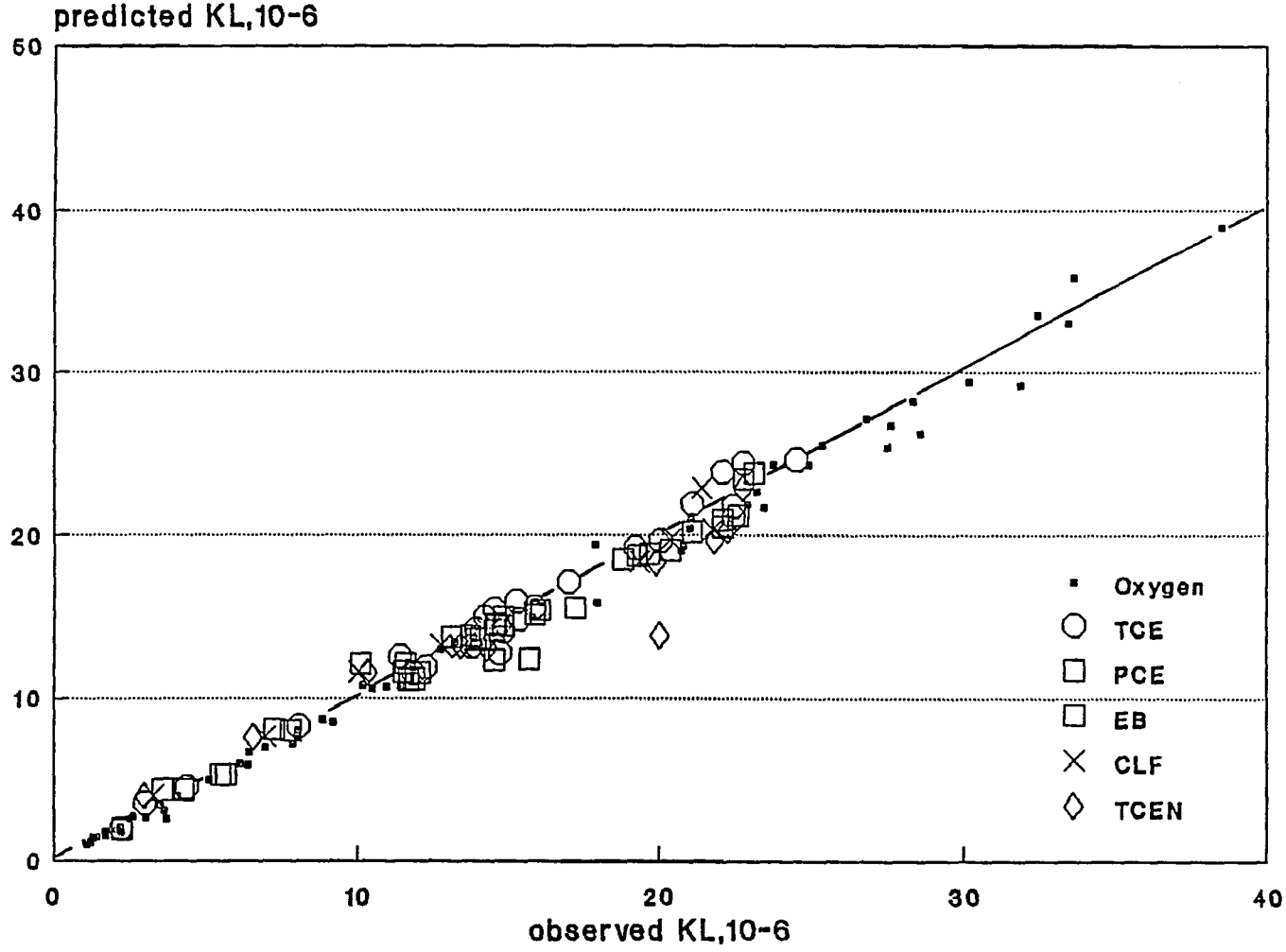


Fig 5.26a Relations among w, HRT, and V
(95% TCE removal, at 1500gpm flow rate)

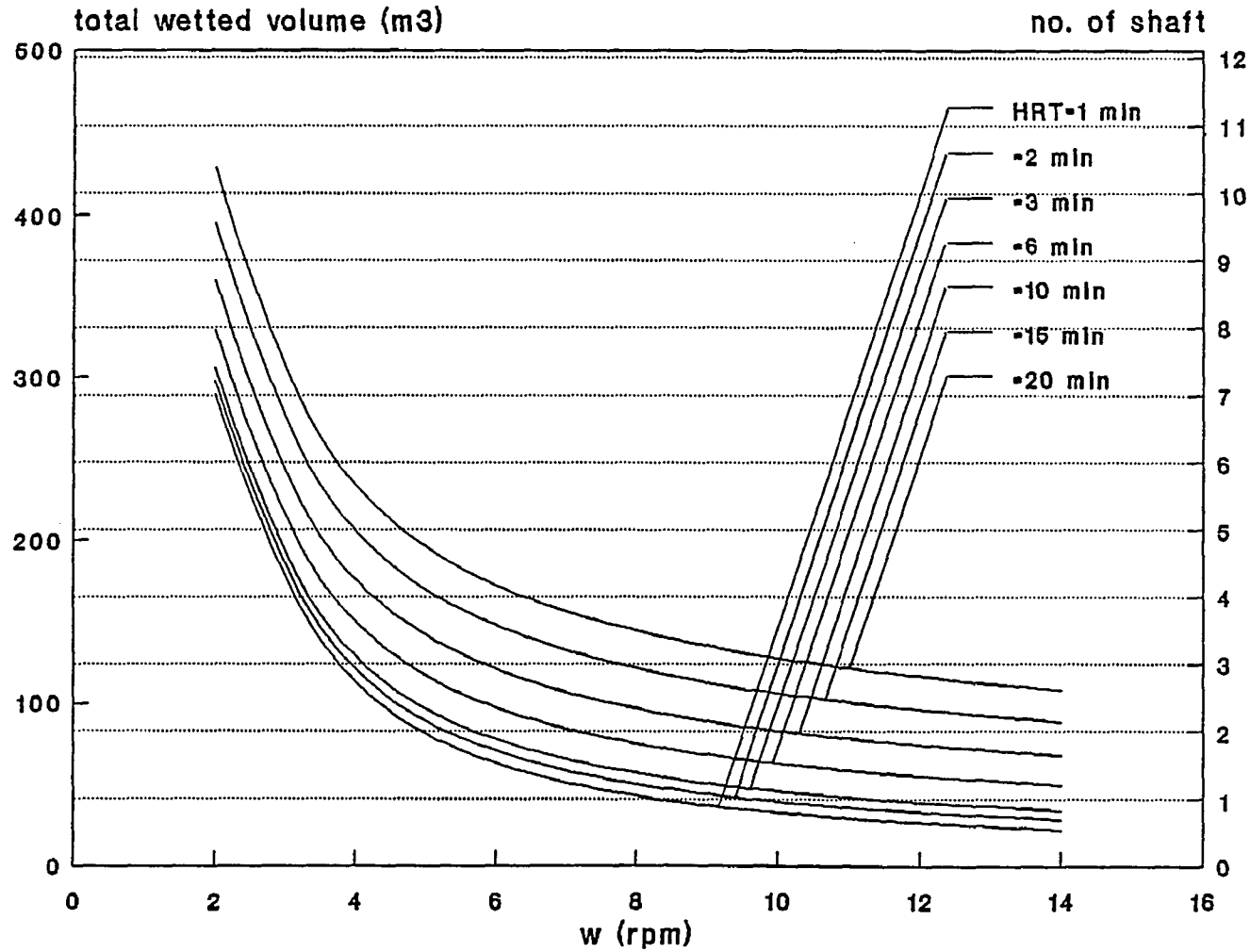


Fig 5.26b Relations among w, HRT, and Energy Requirement)
(95% TCE removal, 1500 gpm infl. rate)

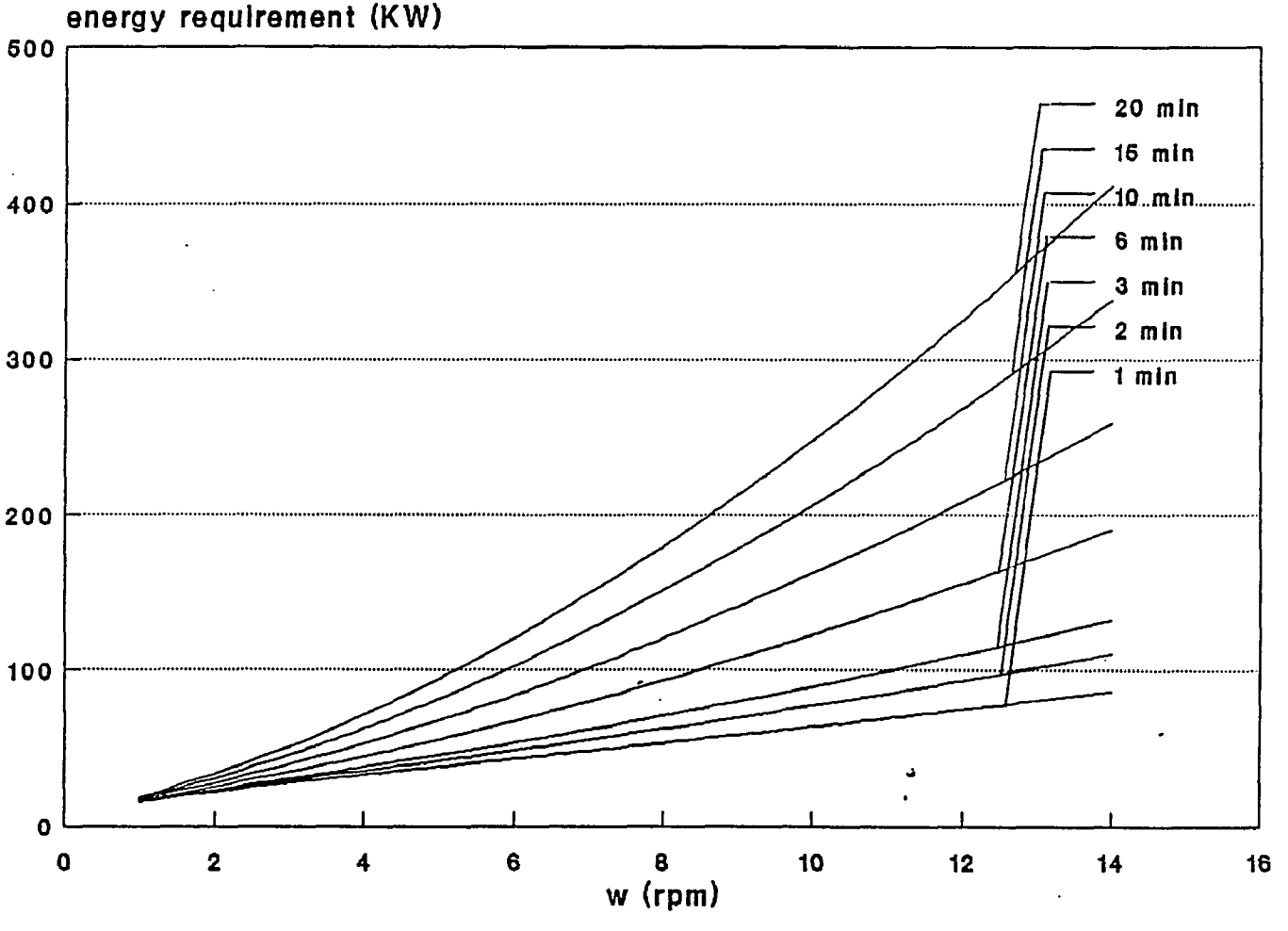


Fig 5.27 Estimated TCE Removal
by Air Stripping in a RBC Unit

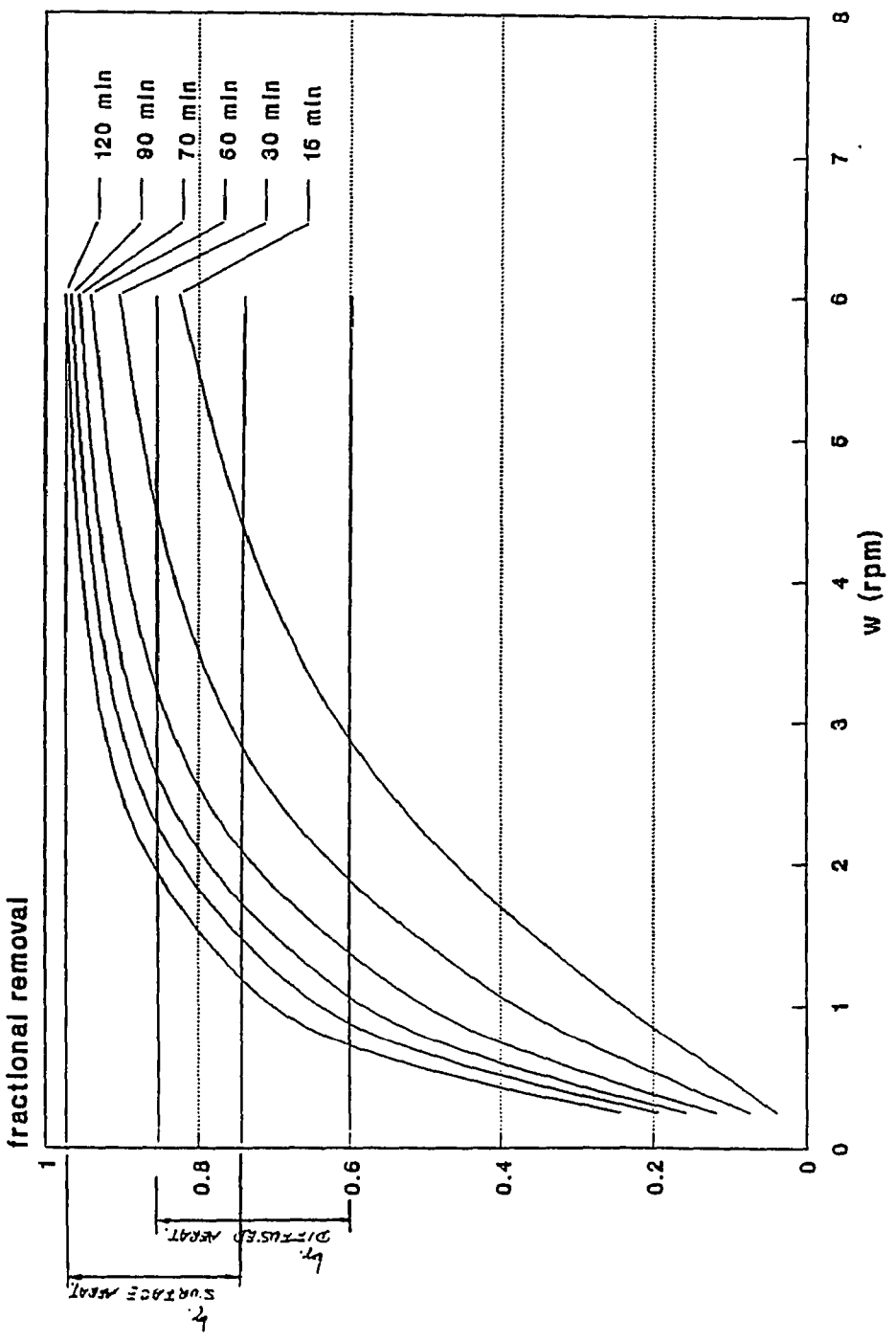
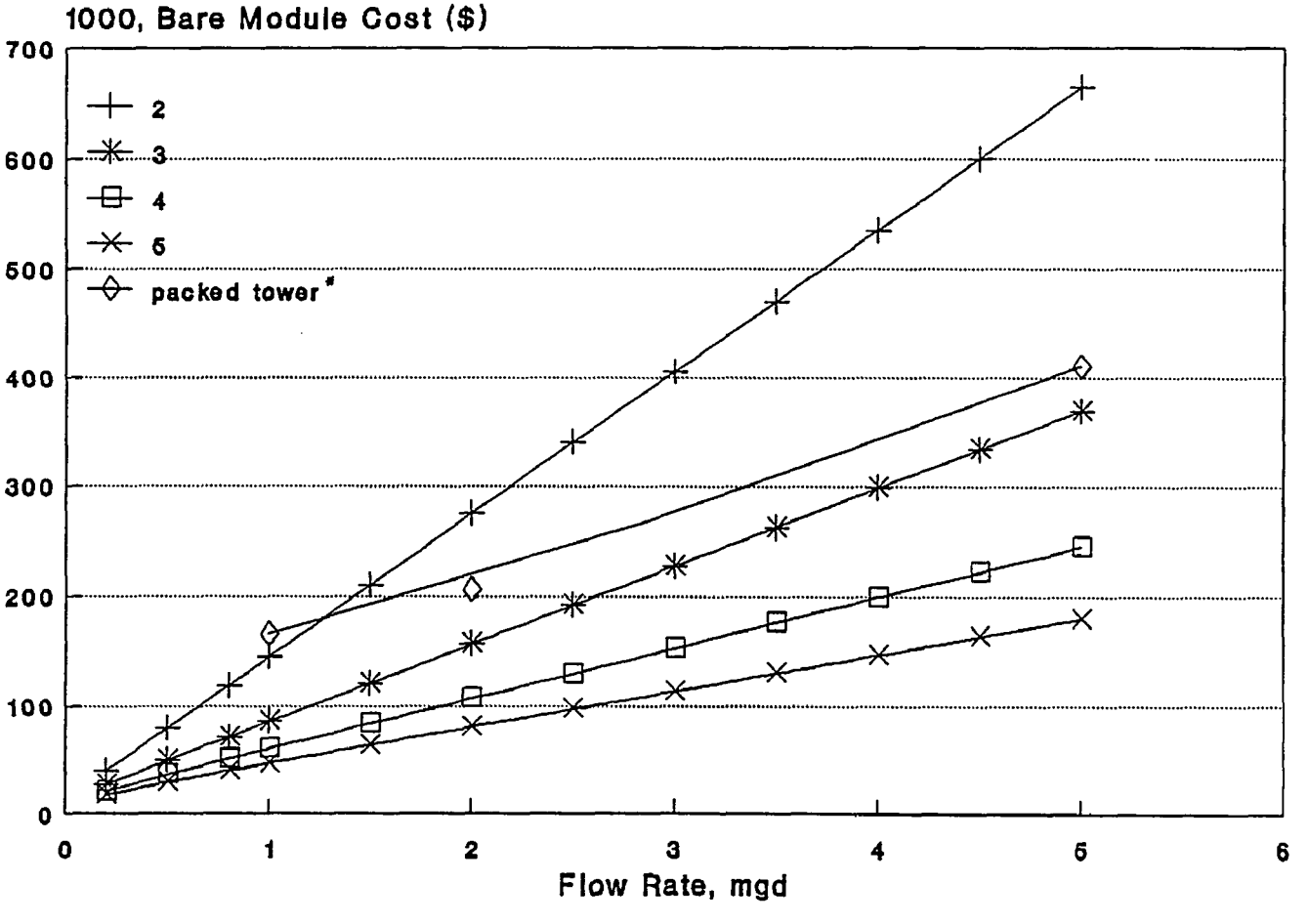


Fig 5.28 Bare Module Cost for RDC and Packed Tower* for 95% TCE Removal
(Field Material/labor+Equipment)



*Data from Ruggiero (1982)

CHAPTER 6 CONCLUSIONS

The literature and the data obtained in this study suggested that a single RDC stage behaves like a continuous flow stirred tank. Without substantial accumulation of the diffusing components in the surrounding air, the absorption/desorption of all six compounds studied in this work could be interpreted satisfactorily by using a model semi-empirically developed on the basis of liquid-phase resistance controlled diffusion process.

The model developed in this study providesa comprehensive correlation between the mass transfer coefficient and operational parameters as shown by Eq. 5.1 to 5.5. Those equations have been proved to be valid for various experimental conditions by the data obtained in this work. In general, both the bulk water surface and the disk surface contribute to the overall mass transfer rate in an RDC system. At higher rotational speeds, the absorption of oxygen or the desorption of VOCs is mainly controlled by the transfer from the water film on the disk surface. However, at lower speeds, the transfer from the agitated water surface becomes dominant. Since the relative importance of the water surface and the disk surface are somewhat dependent upon the RDC systems used and the actual operating conditions, an evaluation should be made before

applying the model to a specific RDC system. For common commercial RDC systems, the data obtained in this study should provide valuable information in estimating the actual mass transfer rates. However, only physical stripping has been considered in this study. With respect to the impact of the biomass on stripping efficiency, a further study is needed.

The rate of mass transfer of VOCs was shown to be dependent on the solute's diffusivity in water to a power approximately equal to 0.5. This exponent coefficient was independent of the rotational speed and other operating conditions. Since the aeration rate is usually available for a specific RDC system, the stripping rates of VOCs can be estimated from the oxygen transfer rate by Eq. 5.6 and 5.8. The data of this study shows that the gas-phase resistance is negligible under higher rotational speeds or for highly volatile components. However, for moderately volatile compounds, chloroform for example, the gas-phase resistance may have been more significant than was predicted by considering liquid phase resistance only. The estimated value of $k_G a/k_L a$ was found to be 45 from this study, which was substantially smaller than the usually assumed value. The influence of the gas phase resistance was evaluated indirectly in this study. Further work may be required to assess the practical significance of gas-phase resistance on the stripping rate.

To achieve the same degree of treatment, the RDC systems entails a greater volume and area than the packed tower. Since the cost per unit volume of media is lower for the RDC, the preliminary economical analysis shows that the capital cost is lower for the RDC system than for the packed tower; if cheap land is available. The energy consumption is somewhat higher for the RDC system. However, the operating and maintenance labor cost is expected to be lower for the RDC system. When there is the risk of inorganic precipitation, growth of biomass, clogging by solids and particles to limit the application of packed tower, RDC can be a good alternative for air stripping.

It has been shown that a great potential exists for the RDC system to enhance biodegrading of the VOCs with the aid of air stripping. Under carefully selected operating conditions, it is possible to decrease the VOC concentration to a level that will not violate the air pollution permit and at the same time, allow biodegradation to occur. The combination of the air stripping process with the biological process, in a RDC system to remove VOCs deserves further study and evaluation for the control of VOCs.

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